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Evaluation of Dry Sorbents and Fabric Filtration for FGD (Flue Gas Desulfurization)

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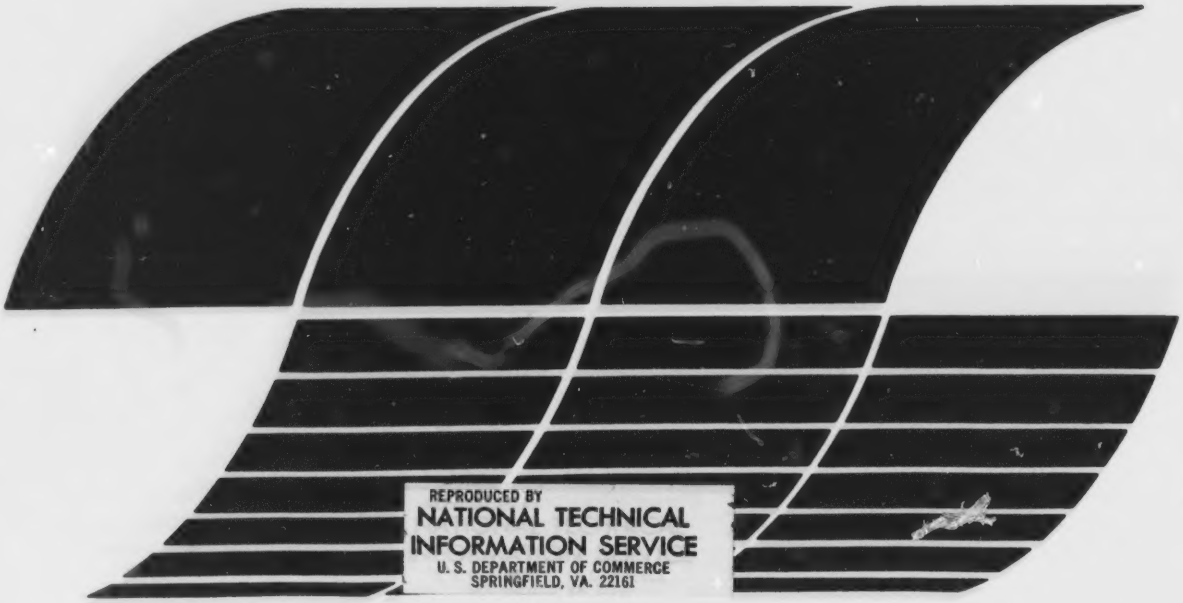

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16. ABSTRACT The report gives results of a study to assess the use of baghouses (fabric filtration) to control air pollutant emissions (particularly SOx) from large utility combustion sources. The assessment included sorbent costs, and system capital, operating, and disposal costs. SO2 would be removed by introducing powdered dry sorbent into the gas stream or by precoating the baghouse fabric with sorbent. The objective of the study was to determine if the apparent economic advantage exhibited by the concept would remain intact after independent third-party evaluation and if the economic (and other) advantages are sufficiently large to warrant further development of the process at field installations. The evaluation shows that the dry sorbent baghouse FGD process exhibits an economic advantage when compared with current lime and limestone scrubbing technology when applied to Western power plants burning low sulfur coal. Further demonstrations on the pilot plant scale have been recommended, particularly at high temperatures. The need for extensive user (electric utility) commitment, in order to justify the considerable capital investment needed to open a commercial-scale sorbent (nahcolite) mine may represent the greatest barrier to commercialization of the process.

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Desulfurization	Coal	Baghouses	07A,07D	21D
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ABSTRACT

The use of conventional baghouses has long been an accepted method for treating exhaust gases from certain industrial and combustion sources. With the advent of more stringent particle control standards and the expanded use of low sulfur coals, interest in baghouses as a control device for large utility combustion sources is growing. In recent years, certain baghouse vendors and natural resource companies having an interest in materials which are potential SO_2 sorbents, have promoted the use of the baghouse as a flue gas desulfurization (FGD) device. SO_2 removal is accomplished by introducing powdered, dry sorbent into the gas stream or by precoating the baghouse fabric with sorbent. In some instances, experimental studies were conducted and reported.

In order to further evaluate the concept prior to embarking on expensive field test work, EPA requested that TRW conduct a more thorough study to include independent assessment of sorbent costs, system capital costs, system operating costs and disposal costs. The basic objective of the present study is to determine whether or not the apparent economic advantage exhibited by the concept remains in tact after independent, third-party evaluation and whether the economic (and other) advantages are sufficiently large to warrant further development of the process at field installations.

The evaluation shows that the dry sorbent baghouse FGD process exhibits an economic advantage when compared with current lime and limestone scrubbing technology when applied to western power plants burning low sulfur coal. Further demonstrations on the pilot plant scale for the dry sorbent baghouse technology have been recommended, particulary at elevated temperatures. The need for substantial user (electric utility) commitment in order to justify the considerable capital investment needed to open a commercial-scale sorbent (nahcolite) mine may represent the greatest barrier to the commercialization of this process.

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SECTION 1

EXECUTIVE SUMMARY

The use of dry chemical sorbents in conjunction with a conventional baghouse to remove sulfur oxides from flue gases has been examined in several test programs conducted by potential users and vendors of baghouses, fabrics, and sorbent materials. These various test programs have served to prove the technical feasibility of the process, but have not examined, in detail, the economic aspects of a dry sorbent/baghouse flue gas desulfurization (FGD) system. In order to provide additional input to further decisions regarding development of the concept, a study was undertaken to estimate costs and evaluate problems associated with a typical commercial installation. By estimating costs using a situation which approximates a "best case" and comparing these costs with a currently available FGD system, it was possible to determine if the dry sorbent/baghouse FGD process represents a substantial breakthrough in FGD technology and can be expected to attain substantial market penetration in the future.

"Best case" conditions were selected after examination of: current trends in boiler design, coal supplies in the vicinity of potential sorbent supplies, SO_2 removal requirements, and available experimental data on dry sorbent performance. The conditions that were selected for detailed economic evaluation were:

- A new 500 MW boiler
- 1% sulfur western coal
- 750 mile (1,200 km) distance from nahcolite supplies
- Semi-arid region with the water table 50 ft (15 m) below the surface

- 400°F Baghouse temperature
- 70% SO_2 removal required

It is important to note that, although additional refinement of conditions may produce a slightly lower-cost base-case for this study, any conditions which are site specific or narrow in application would be unrealistic since the equivalent of ten 500 MW installations would be required to support one commercial-scale nahcolite mine. This is approximately 4% of the power output of the committed fossil-fueled generating plants for the period 1977-1986.

SORBENT SUPPLY AND COSTS

Data indicate that nahcolite (raw $NaHCO_3$) is the only practical material for consideration in the dry sorbent/baghouse FGD process. The total recoverable reserves of nahcolite are large enough to support even the most widespread application of the process and can be mined at reasonable cost. It is most important, however, to note that nahcolite is not currently mined and that the capital investment involved in opening a commercial-scale mine is substantial. A minimum-sized (1,000,000 tons/year) commercial mine would require the nahcolite demand of ten 500 MW installations each requiring 1,000,000 tons/year as described in the base-case examined in this study. Thus, the major obstacle to the dry sorbent/baghouse FGD concept from the standpoint of sorbent supply is the need for substantial commitment from potential users (5,000 MW) in order to open a mine. In order to obtain such a large commitment to an untried concept, it would be necessary for the concept to exhibit very large commercial economic benefit over current technology for which supplies of sorbent (lime and limestone) are readily available.

WASTE DISPOSAL

A number of concepts for disposal of the mixed fly ash/sodium salt waste material generated by the proposed process have been examined. These include: landfill, ocean disposal, mine disposal, deep welling, brine ponds, insolubilization, and use in chemical processes. Of the concepts examined, landfill appears to be both generally applicable and reasonably inexpensive. A detailed design and discussion of operation of a landfill designed to provide adequate environmental controls for the highly soluble waste product involved are included in this report. The basic concept of the landfill design involves protecting the placed material from intrusion by groundwater, surface water and rainwater in order to prevent leaching, rather than attempting to contain leachate, as is necessary with disposal of wet materials such as scrubber sludge. The existence of vast natural deposits of soluble materials such as sodium chloride, trona and nahcolite indicates that conditions can exist which protect these from water intrusion and thus enable storage for very long periods of time. A cost estimate of $6 per ton of material placed is detailed in Section 5.

ECONOMICS

In order to determine if the dry sorbent/baghouse process can be expected to achieve commercial market penetration upon successful completion of a development program, an independent cost estimate for a typical commercial installation was prepared. To the extent practical, estimating procedures consistent with those used by TVA(2) in their costing of commercial FGD processes were used. Independent estimates for sorbent supply and waste disposal were prepared and all costs were adjusted to a 1977 basis. The valid comparison of the dry sorbent/baghouse FGD system to the limestone scrubbing process required that both desulfurization processes use the same physical conditions and inputs (i.e., both were located at a site in Wyoming burning 1% sulfur coal with an ash content of 10%). Also, both included particle control capable of meeting the Federal New Source Performance Standard of 0.1 lb. particulate matter emitted per million BTU's of heat input.

Using capital costs as a basis of comparison, the dry sorbent/baghouse system is significantly lower in cost than the limestone scrubbing system, even taking into account the accuracy of the cost estimating procedures developed by TVA. Using the capital costs derived in this report, the dry sorbent/baghouse FGD system is seen to be slightly less than 40 percent of the costs for a limestone scrubber serving a similar power unit. The differences are not so great when comparing the annualized operating costs of the two systems; in this comparison, the dry sorbent/baghouse system is estimated to be approximately 35 percent less than the limestone scrubber. However, in view of the accuracy of the cost estimations, this difference may or may not be significant. Nonetheless, the dry sorbent/baghouse does show promise as a viable alternative to the present state-of-the-art in flue gas desulfurization.

The possibility of developing a regenerable dry sorbent/baghouse process using magnesium oxide as the sorbent was examined and is discussed in Appendix D. Kinetic data show that such a system is not feasible.

RECOMMENDATIONS

The dry sorbent/baghouse flue gas desulfurization process appears to demonstrate an economic advantage over other currently available technologies for flue gas desulfurization, when applied to western power plants burning low sulfur coal as fuel. Further demonstrations of this process should be pursued. A pilot plant study appears warranted to provide confirming technical and cost data.

SECTION 2

INTRODUCTION

The use of conventional baghouses, employing fabrics for filtration of particulate matter, has long been an accepted method for treating exhaust gases from certain industrial and combustion sources. With the advent of more stringent particle control standards and the expanded use of low sulfur coals, interest in baghouses as a control device for large utility combustion sources is growing. This is due to ever increasing costs for electrostatic precipitators which are not well suited for extremely high efficiency operation with low sulfur flue gases. In recent years, certain baghouse vendors and natural resource companies having an interest in materials which are potential SO_2 sorbents, have promoted the use of the baghouse as a flue gas desulfurization (FGD) device. SO_2 removal is accomplished by introducing powdered, dry sorbent into the gas stream or by precoating the baghouse fabric with sorbent. In some instances, experimental studies were conducted and reported.

In the Spring of 1976, a short study of the dry sorbent/baghouse FGD concept was undertaken by TRW, under contract to EPA's Industrial Environmental Research Laboratory at Research Triangle Park, North Carolina. The objectives of this study included(1):

- Gather and evaluate available data and reports concerning the concept.
- Qualitatively evaluate the range of conditions and situations for which the concept is practical.
- Discuss the state of development of baghouses, fabrics, and operating modes as they impact the dry sorbent concept.

- Examine sorbent supply, waste disposal and possible regeneration.
- Define problems and knowledge gaps concerning the concept.
- Recommend future work necessary to generate information to completely assess prospects for commercialization, range of applicability, and impact of the concept.

At approximately the same time, the Electric Power Research Institute (EPRI) contracted with Bechtel Corporation for a study with very similar objectives(8).

Both the TRW/EPA study and the Bechtel/EPRI studies concluded that, based on available information, the dry sorbent/baghouse FGD concept holds considerable promise of economic advantage under certain geographic, process and regulatory constraints. The optimistic view of the concept must, however, be tempered by the fact that experimental data are sketchy, certain major institutional and financial barriers exist, and, most significantly, the available cost estimates were prepared and presented by firms having a potential financial benefit in the eventual commercialization of the process.

In order to further evaluate the concept prior to embarking on expensive field test work, EPA requested that TRW conduct a more thorough study to include independent assessment of sorbent costs, system capital costs, system operating costs and disposal costs. The basic objective of the present study is to determine whether or not the apparent economic advantage exhibited by the concept remains intact after independent, third-party evaluation and whether the economic (and other) advantages are sufficiently large to warrant further development of the process.

SECTION 3

CONCLUSIONS

Published experimental data (8) indicate that nahcolite (impure $NaHCO_3$) is the most promising sorbent for use in the dry sorbent baghouse flue gas desulfurization (FGD) concept. Available data are adequate for cost estimation that is on the order of -20 percent to +50 percent accurate. These data are not, however, sufficient to permit potential users (electric utility companies) to accept commercial bids with confidence. A recent and rather extensive pilot test conducted by Wheelabrator-Frye, Inc., Superior Oil Co., and a midwestern utility consortium may well provide sufficient data upon which to base a commercial design. The extent to which these data will be made public is not known at this time.

The reserves of nahcolite are adequate to supply even very extensive application of the dry sorbent/baghouse FGD concept to western and midwestern low sulfur coal-fired boilers. The material is not currently mined. A reasonable commercial price for nahcolite (70 percent assay) is $25 per ton at the mine mouth and $32.50 per ton at a power plant located approximately 750 miles (1,200 kilometers) from the mine site (northwestern Colorado). IRI and Superior Oil have confirmed that a firm commitment for supplying nahcolite to several large baghouse FGD installations would be necessary before any of the three nahcolite lease holders would be able to invest the significant sums of money required to open a nahcolite production mine.

The wastes generated by a 500 MW application of the dry sorbent/baghouse FGD process can be disposed of in an environmentally acceptable manner. The basic disposal concept involves a landfill designed specifically to preclude groundwater, surface water and rainfall from entering the deposit of soluble waste material. The prevention of groundwater intrusion is accomplished through careful site selection and exclusion of surface water and rainfall by combination of dikes and an impermeable (butyl rubber) membrane placed over the landfill site. Landfill costs for the concept and design described in this report are estimated at $6 per ton of waste.

The dry sorbent/baghouse FGD process, when costed in the base case selected for analysis in this study (500 MW, new installation, 1 percent sulfur coal, 400°F baghouse temperature), shows capital costs of $46 per KW and an annualized operating cost of 2.60 mills per kWh. This is 35% lower than the base case limestone FGD process using those inputs which give a flue gas composition of 4.3×10^6 lb/hr flow, 36,550 lb/hr ash, 8.17×10^3 lb/hr SO_2 and 3.8×10^5 lb/hr moisture, equivalent to the nahcolite/baghouse FGD system. The TVA cost estimates were computed to be $119 per kW and 4.08 mills per kWh for the capital costs and annualized operating costs, respectively. Even in view of the accuracy of the cost estimating methods (-20% to +50%), the cost differences between the dry sorbent/baghouse FGD system and the limestone scrubbing FGD process must be considered promising. This, in conjunction with the fact that the limestone scrubber is considered one of the more economical of flue gas desulfurization schemes, makes the dry sorbent/baghouse system seem quite favorable.

However, since the proposed concept is extremely sensitive to the amount of sulfur removed (more than 40 percent of the annualized operating costs for the dry sorbent/baghouse system is for sorbent and waste disposal compared with 3% for the limestone scrubbing process), applications to high sulfur coal-firing may not be economically competitive with the limestone scrubber.

Experimental data (9, 10) indicate that nahcolite utilization may approach 100% if the dry sorbent baghouse is operated at about 525°F (274°C). This potential improvement in utilization would decrease the costs associated with sorbent and disposal by about 30 percent over the base case studied in this report. Since these costs represent slightly less than 45 percent of the total annual cost, a potential savings of nearly 15 percent is possible. This savings would be somewhat offset by increased capital costs associated with the higher flue gas temperatures. The net effect should amount to an approximate 10 percent reduction in the base cost of 2.60 mills per kWh. The new cost of 2.34 mills per kWh is lower than the limestone scrubbing cost of 4.08 mills per kWh by approximately 40 percent.

The use of MgO as a dry sorbent in a regeneration scheme was briefly examined. Kinetic data developed in Appendix D show that such a system, in conjunction with the baghouse FGD concept is not feasible.

SECTION 4

RECOMMENDATIONS

The dry sorbent/baghouse flue gas desulfurization process appears to demonstrate an economic advantage over other currently available technologies for flue gas desulfurization, when applied to western power plants burning low sulfur coal as fuel. Thus, further demonstrations of the dry sorbent FGD process utilizing nahcolite as a sorbent should be pursued. Since the results of this analysis demonstrate that the dry sorbent process FGD system becomes economically advantageous when operated at temperatures in excess of those used for current demonstrations, a pilot plant study appears warranted to provide confirming technical and cost data for this study.

Since operation at extreme elevated temperatures of approximately 525°F (274°C) holds some additional promise for cost reduction due to increased utilization of the sorbent, the development of better test data at these temperatures should be pursued. The use of existing test facilities may be a possible route to obtain technical data in this area.

The data from tests currently in operation with Wheelabrator Frye, Inc., and Superior Oil Co., should be examined when they become available. Results from these tests should be used to confirm data obtained from other sources in developing the assumptions used in preparing this report. Additionally, these tests may provide some additional data on nahcolite utilization rates that can be expected at elevated temperatures.

SECTION 5

DISCUSSION

SYSTEM DESIGN

The applicability of dry sorbent baghouse as an acceptable means of flue gas desulfurization (FGD) is dependent on many technical and environmental considerations. The system design as developed in this report has resulted from a systematic evaluation of these considerations, and represents a first order optimized design for the dry sorbent/baghouse FGD system as applied to those power plants for which this type of FGD system is most applicable.

Generating Station Design

The design parameters of the power generating station have been chosen to represent those found at typical modern plants. Wherever possible, the parameters used by the Tennessee Valley Authority in their report Detailed Cost Estimates for Advanced Effluent Desulfurization Processes(2) are adopted in this design. This design consistency allows the process design data and economic criteria developed in this report to be directly compared with those found for the other major flue gas desulfurization processes located in the southwestern United States.

Type of Plant--

A new plant installation of the dry sorbent/baghouse FGD system is expected to yield the most beneficial results. The plant operating parameters can be established on the basis of the overall installation rather than on only the power generation section, thus resulting in an overall increase in plant efficiency. Installation costs are minimized and costly modification of existing structures and equipment are eliminated. Other benefits of a baghouse operation can be considered in the evaluation of plant design.

Choice of Fuels--There are three basic types of fossil fuels used for power production: oil, gas, and coal. Of these three, natural gas is in increasingly short supply and few if any new plants will be relying on this form of energy. The use of oil for power production is common in certain coastal and oil producing regions of the United States, but the future supply of this energy source is becoming increasingly doubtful for long term utilization. Coal, on the other hand, is plentiful in many parts of the United States, thus assuring its long-term availability.

Coal has an additional property which makes the dry sorbent baghouse advantageous over other FGD systems. Particle collection to provide opacity control has become increasingly important as Federal and state regulations for this parameter grow increasingly more stringent. This is particularly true for power plants burning low sulfur western coal. The resistivity of the fly ash produced by the combustion of this coal makes the particles exceedingly difficult to collect by normal methods, e.g., electrostatic precipitators. Fortunately, the baghouses supplied for dry sorbent FGD will also provide exceptional particle control. Baghouses installed in fossil-fired power plants typically provide a particle collection efficiency greater than 99.5 percent. Figures 1 and 2 present collection efficiency data obtained by test at the NUCLA Generating Plant and at Sunbury Steam-Electric Station, respectively.

Fuel Transportation--It is assumed for the purpose of this study that coal is delivered to the generating station by railroad, where it is unloaded and stored in piles for use. This is the most common method of coal transportation, although other methods have been utilized in specific cases. It has therefore been assumed that the existing rail spur would be available to handle the transportation of the dry sorbent to the power plant.

Size--The major capital expenses involved in the construction of a dry sorbent baghouse are relatively insensitive to size. These are the fresh sorbent unloading facilities, the design engineering expenses, and to a limited extent the baghouse itself. It is therefore important to select an appropriate station size for consideration in this study. Table 1 summarizes all presently committed new fossil fueled generating capacity for the next

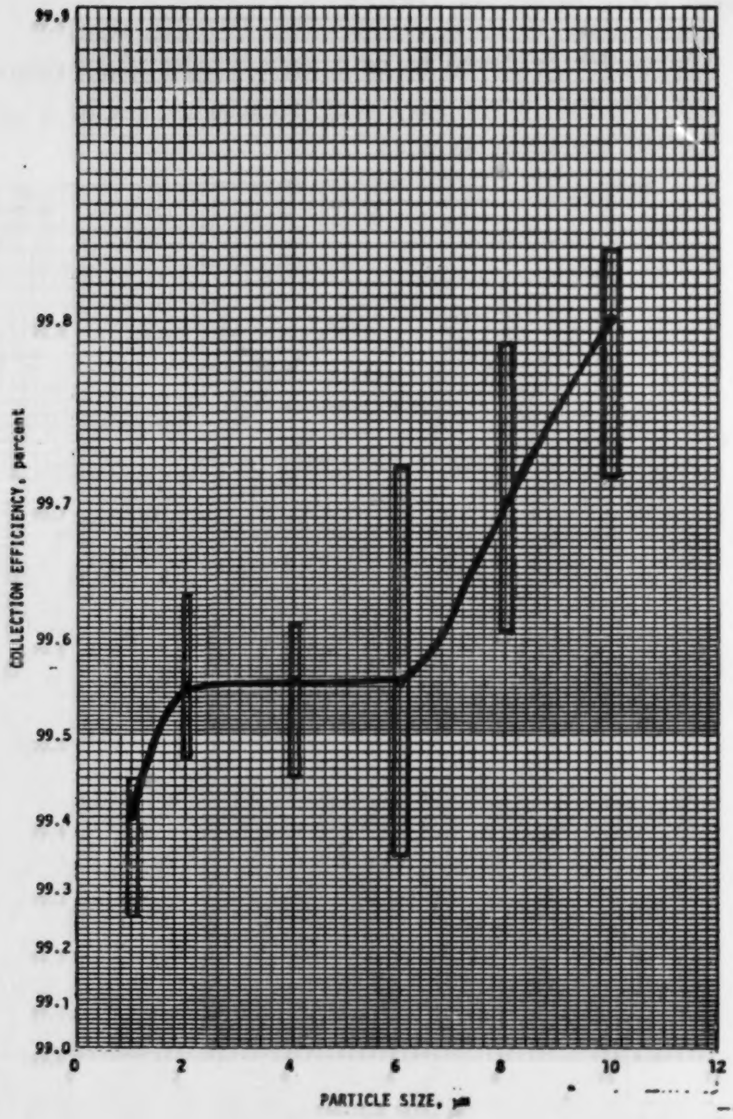


FIGURE 1. COLLECTION EFFICIENCY OF BAGHOUSE AT NUCLA GENERATING PLANT(3)

A. Curve based on average of all runs except those with one set of substrates and run 16

B. Curve based on average of all used bag runs

Note: Curves are believed to be biased toward lower removal efficiency for 8μm and larger particles because of use of a cyclone percollector on the inlet samples.

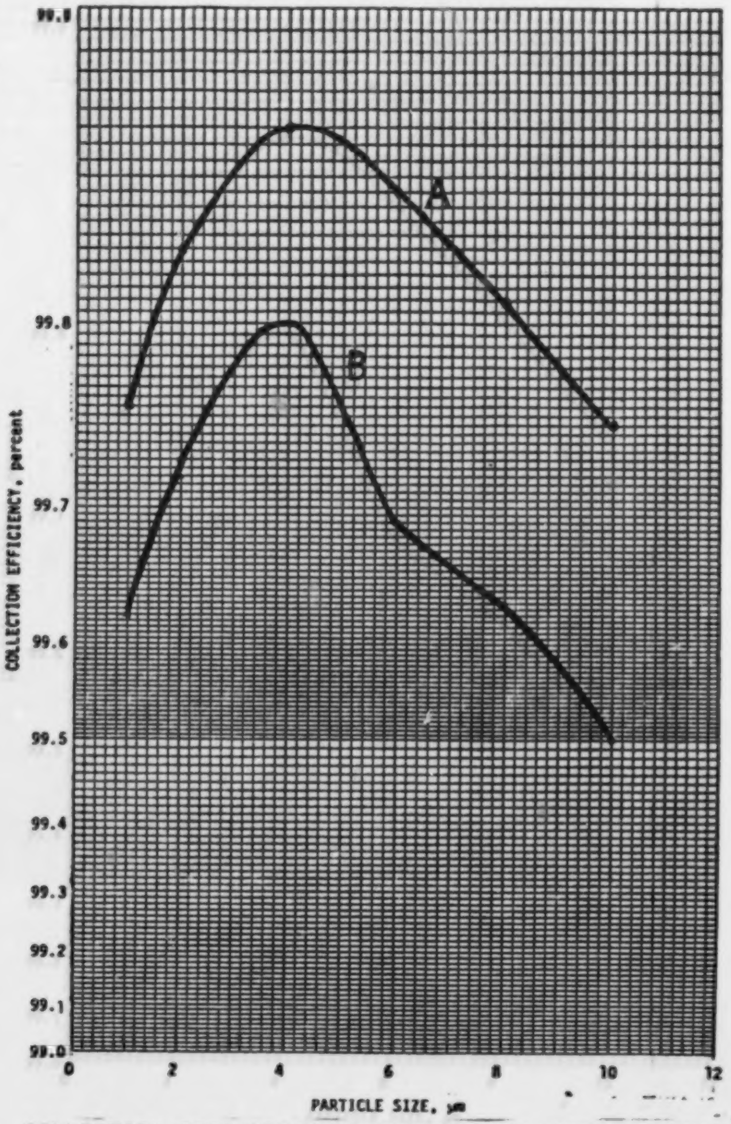


FIGURE 2. COLLECTION EFFICIENCY OF BAGHOUSE AT SUNBURY STEAM-ELECTRIC STATION(4)

10 years. From these figures it is readily apparent that the average size of all new generating units is 500 megawatts. This has therefore been selected as the appropriate plant size on which to base this design estimate.

TABLE 1. COMMITTED FOSSIL FUELED GENERATING PLANTS 1977-1986(5)

Year	Number of Plants	Combined MW Capacity	Average MW Capacity
1977	37	17,160	464
1978	34	14,424	424
1979	37	16,656	450
1980	31	15,421	497
1981	32	14,515	454
1982	25	13,732	549
1983	24	12,959	540
1984	20	11,307	565
1985	21	11,770	560
1986	9	4,688	521
Total	270	132,632	491

Location of Plant--

The dry sorbent baghouse FGD system requires a considerable amount of raw materials for operation. Because of this, the transportation of the sorbent material becomes a significant factor in the operating costs of the facility. Additional constraints are placed on the choice of plant location because of particularities of the waste disposal system which must necessarily be designed for each specific site.

Region--We have limited our review of plant locations to an area within approximately 750 miles (1200 kilometers) of the source of sorbent supply. This area includes most or all of the following states: Arizona, California, Colorado, Idaho, Kansas, Montana, Nebraska, Nevada, New Mexico, North Dakota, Oklahoma, Oregon, South Dakota, Texas, Utah, Washington, and Wyoming. This does not imply that dry sorbent baghouses are not practical in other regions; however the operating costs incurred for this type of system in such a region would certainly be greater than for the considered situation.

Local Geology--The power plant is situated on level ground suitable for industrial construction without excessive foundation and ground preparation work. Rainfall is assumed to be average for an arid area typical of the southwest United States.

The waste disposal location is assumed to be approximately 15 miles (24 kilometers) from the power plant. It has a flat terrain, predominantly composed of clay soils. The water table is greater than 100 feet (30.5 meters) below surface area; there is no perched water table; and there are no surface streams in the immediate area.

Coal--

The operating costs and removal efficiency of the dry sorbent FGD process are dependent on ash content, sulfur content, and heating value of the coal. These coal parameters have a considerable effect on the design and sizing of the specific components. It is therefore necessary to assign specific parameters to the assumed coal supply to evaluate the capital costs of such an installation.

Coal Availability--Coal deposits are plentiful in the southwest region which is considered for the plant location. The actual supply for specific plants is assumed to be mined locally. Figure 3 shows the major coal deposits(5) within a 750 mile (1200 kilometers) radius of northwest Colorado, which is the assumed location of the sorbent supply.

Coal Analysis--A comparative analysis of each of the available coal deposits is illustrated in Table 2. The sulfur content of the specific coal used will affect the amount of sorbent necessary to achieve a given SO_2 removal from the flue gas. The sulfur content of the available coals range from 0.27 to 1.3 percent. The ash content of the specific coal used will determine the amount of material collected in the baghouse, thus affecting the baghouse sizing, waste storage sizing, and waste disposal operations. The ash content of the available coals ranges from 3.4 to 13.8 percent. The heating value of the coal used will determine the amount of coal which is necessarily combusted to produce a given heat output from the power plant. An increased coal heating value will decrease both the amount of fly ash (for a given power output) produced and the rate of sorbent consumption. The heating values of the

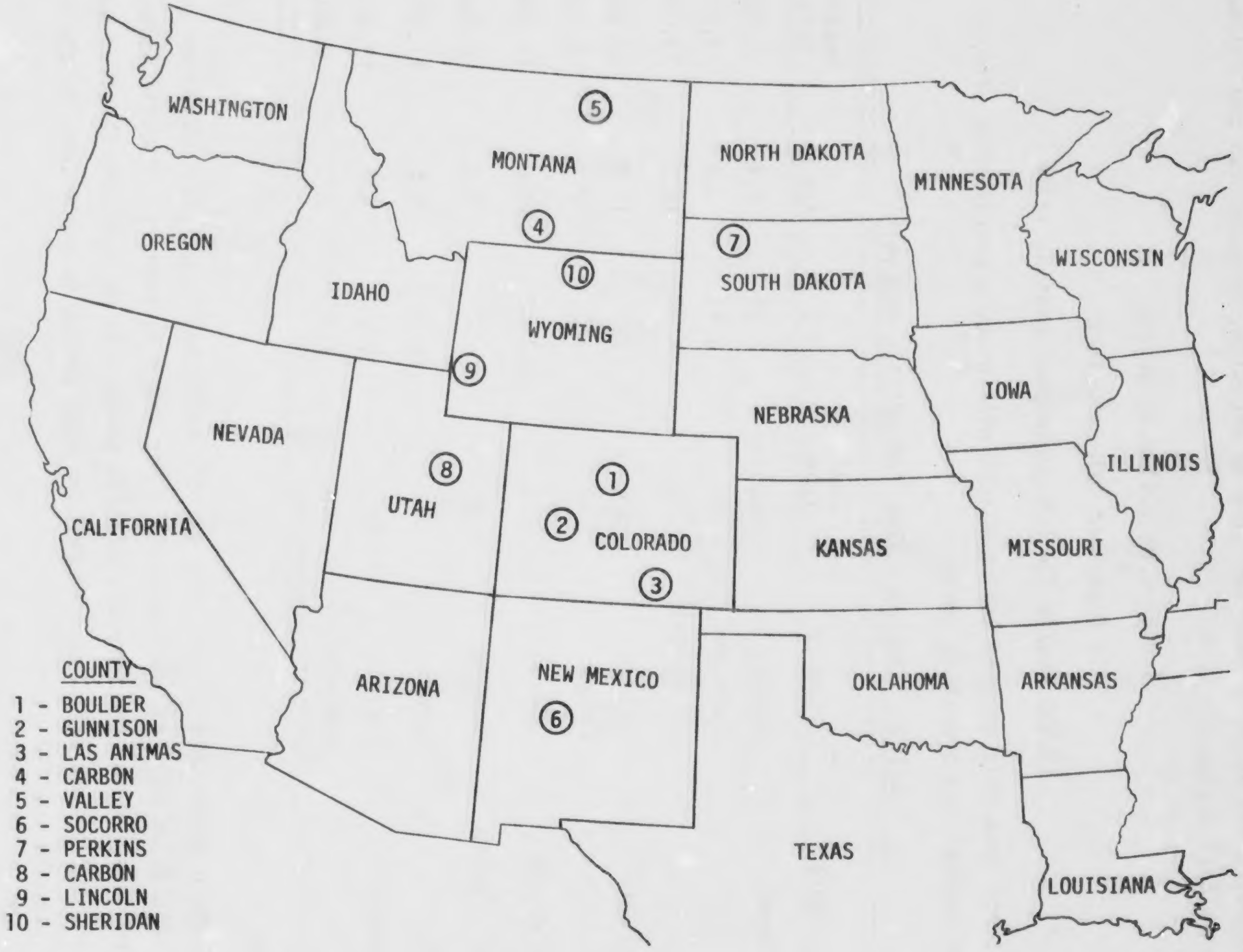


FIGURE 3. MAJOR COAL DEPOSITS WITHIN 750 MILE RADIUS OF NAHCOLITE MINE

available coals range from 6,000 to 13,500 Btu/lb (14,000 to 31,400 joules/gm) For the base case used in this study each of these three parameters have been evaluated, based on the number and size of deposits, to establish the following norm coal analysis:

1.0 percent sulfur by weight

10.0 percent ash by weight

10,500 Btu/lb (24,000 joules/gm) heating value

Variations from this assumed coal analysis will produce predictable effects on the capital and operating costs.

TABLE 2. AVAILABLE COAL COMPARATIVE ANALYSIS(6)

State	County	Coal Type	S %	Ash %	Heating Value
Colorado	Boulder	Subbit.	0.27	5.4	10,000
	Gunnison	Bit.	0.43	3.4	13,500
	Las Animas	Bit.	0.70	12.8	13,000
Montana	Carbon	Subbit.	1.1	11.2	10,500
	Valley	Lig.	1.3	9.1	6,700
New Mexico	Socorro	Bit.	0.82	13.8	12,300
South Dakota	Perkins	Lig.	1.2	9.0	6,000
Utah	Carbon	Bit.	0.6	5.6	12,500
Wyoming	Lincoln	Bit.	1.0	5.5	13,300
	Sheridan	Subbit.	1.1	7.9	9,000

Flue Gas--

Certain parameters of the flue gas are sensitive to utility operation and boiler design as well as coal analysis. For these parameters, specific values have been assumed for the purposes of this study. In particular, the flue gas flow rate and boiler exit temperature have been chosen to correspond to these values used in the TVA study(2):

4.3×10^6 lb/hr (1.95×10^6 kg/hr) mass flow rate
8.9×10^5 scfm (1.51×10^6 m^3/hr)
890°F (477°C) boiler exit temperature

Chemical analysis of the flue gas at boiler exit is assumed to contain the constituents listed below. The proportions of each constituent are considered to be typical of western coal-firing(7).

N_2 - 74%
O_2 - 4.8%
CO_2 - 12.3%
SO_2 - Variable
SO_3 - 1.0% of the SO_2 value
NO - .06%
HCl - .01%
H_2O - 8.8%

Flue Gas Desulfurization Requirements--

The amount of SO_2 removal required for plant operation varies with plant location and coal analysis. Because of the large amounts of sorbent required for a full-size commercial plant, the sorbent handling system design, site storage facilities, and baghouse sizing are affected by this parameter.

Federal and State Regulations--The Federal New Source Performance Standards concerning the SO_2 emissions from a fossil-fired power generating station are currently under revision. Although the new regulations are expected to require 85% removal for most of the available coal types, these anticipated regulations have not been used for this report due to their uncertainty. The existing regulations are based on an allowable emission rate which is proportional to the power output of the plant. This regulation is 1.2 lbs of SO_2 per million Btu (5.16×10^{-7} g/joule) heat input to the boiler. Several states in the southwest have set considerably more stringent regulations for SO_2 emissions. Most of these are consistent with the Federal method and only differ in the amount of SO_2 emissions that are allowable. Utah has set a regulation requiring 80% removal independent of plant size or fuel analysis. Table 3 illustrates the various regulations currently affecting these power plants.

TABLE 3. SO_2 EMISSION REGULATIONS FOR SELECTED STATES

State	Federal Emission Requirements (lb SO_2 per 10^6 Btu Heat Input)	State Emission Requirements (lb SO_2 per 10^6 Btu Heat Input)
Arizona	1.2	0.8
Colorado	1.2	0.3 or 150 ppm
Kansas	1.2	1.2
Montana	1.2	1.2
Nebraska	1.2	1.2
Nevada	1.2	0.21
New Mexico	1.2	0.34
North Dakota	1.2	1.2
South Dakota	1.2	1.2
Texas	1.2	1.2
Utah	1.2	80% Removal
Wyoming	1.2	0.2

Removal Requirements by Location--The most common situation for southwestern power plants will be for a plant to purchase fuel from the nearest coal source. A good estimation of removal requirements is then to evaluate each of the coal deposits in Table 2 with the assumption that a power plant is situated in close proximity to the deposit, and within the same state. The percent SO_2 removal for each of the deposit locations are as follows:

Boulder - 48%
Gunnison - 47%
Las Animas - 71%
Carbon - 40%
Valley - 68%
Socorro - 73%
Perkins - 69%
Carbon - 80%
Lincoln - 86%
Sheridan - 91%

Base Design Removal Assumptions--It can be seen from the above listing of SO_2 removal requirements that 3 of the deposits require little SO_2 removal capability (approximately 40-50 percent), 3 require extensive removal capability (approximately 80-90 percent), and the remaining 4 deposits require a moderate removal capacity (65-75 percent). Based on these data, 70 percent SO_2 removal has been used as the baseline removal requirement for this study.

Sorbent Supply--

Nahcolite is the principal sorbent under consideration for FGD processes. It is found in conjunction with oil shale deposits in the Green River formation of northwestern Colorado. This report assumes a typical assay of nahcolite ore of 70 percent sodium bicarbonate.

Process Design

The process design contained herein represents an estimation of full load operating parameters calculated for the generating station design described in the preceding section. Combustion parameters are included for coal consumption, combustion air requirements, and particular flue gas characteristics. Process design, from a functional standpoint, is presented for a plant using a baghouse flue gas desulfurization system as depicted in this report. This design is based on a careful review of previous test work with dry sorbent flue gas desulfurization to ascertain the effect of various design manipulations on overall plant effectiveness. From this analysis, a particular operating plan has been established to make full use of the dry sorbent capabilities and to minimize operating costs.

Coal Combustion--

The process of coal combustion to produce heat for steam generation is a normal industrial operation, occurring with great frequency throughout the world. No radical or improved designs are expected in this technology, and none are considered in this report. The combustion calculations included in this report are based on historical utility boiler operations and are presented to display the basic operating parameters upon which the flue gas desulfurization system design has been based.

Coal Requirements--The Tennessee Valley Authority has established the basic heat rate requirement for their operating system(2). The fuel Btu input requirements to produce a fixed power generation is a function of the unit size and time of construction. This relationship can be stated as:

Power Unit Input Heat Requirement

Size, MW	Status	Heat Rate, Btu/kWh
1,000	New	8,700
1,000	Existing	9,000
500	New	9,000
500	Existing	9,200
200	New	9,200
200	Existing	9,500

The design of a new 500 MW utility boiler may therefore assume a heat rate of 9,000 Btu/kWh (9,500 kjoules/kWh). The quantity of coal required to produce 500 MW of electricity, assuming a coal heat value of 10,500 Btu/lb (24,400 joules/gram) is:

$$Q_{lb/hr} = (500\ \text{MW})\ (9{,}000\ \text{Btu/kWh})\ (1{,}000\ \text{kW/MW})\ \left(\frac{1}{10{,}500\ \text{Btu/lb}}\right) \qquad (1)$$

$$= 4.3 \times 10^5\ \text{lb/hr}\ (2.0 \times 10^5\ \text{kg/hr})$$

Fly Ash Production--The coal assumptions included an ash content of 10 percent by weight. A review of typical utility operations demonstrate that approximately 15 percent of the ash content of the coal becomes bottom ash and is not carried by the flue gas stream. The balance, or 85 percent, is assumed to produce fly ash. The amount of fly ash is therefore:

$$Q_{lb/hr} = (4.3 \times 10^5\ \text{lb/hr})\ (0.85)\ (0.10) \qquad (2)$$

$$= 36{,}550\ \text{lb/hr}\ (16{,}575\ \text{kg/hr})$$

Fly ash loading is another parameter required for flue gas desulfurization system design. It is stated in terms of grains per standard cubic foot (gr/scf) and is calculated as follows:

$$\text{Loading} = (36{,}550\ \text{lb/hr})\ (6{,}998\ \text{gr/lb}) \times (0.01667\ \text{hr/min})\ \left(\frac{1}{8.9 \times 10^5\ \text{scf/min}}\right) \tag{3}$$

$$= 4.8\ \text{gr/scf (dry)}\ (10.9\ \text{grams/m}^3)\ \text{(at air preheater entrance)}$$

Sulfur Dioxide Content of Flue Gas--Results of tests performed on utility boilers(6) demonstrate that approximately 95% of the sulfur content of the as-fired coal becomes SO_2 in the flue gas, with the balance of the sulfur remaining in the ash. The molecular weight of sulfur is 32 and oxygen is 16. One pound of sulfur can therefore produce 2 lbs of SO_2. The basic coal analysis assumed a sulfur content of the coal of 1.0% by weight. The rate of SO_2 emission is therefore given by:

$$Q_{\text{lb/hr}} = (4.3 \times 10^5\ \text{lb/hr})\ (0.01)\ (0.95)\ (2) \tag{4}$$

$$= 8.17 \times 10^3\ \text{lb/hr}\ (3.7 \times 10^3\ \text{kg/hr})\ SO_2$$

Since the molecular weight of SO_2 is 64, the pound-moles (lb-mole) of SO_2 produced is:

$$Q_{\text{lb-mole/hr}} = (8.17 \times 10^3\ \text{lb/hr})\ \left(\frac{1}{64\ \text{lb/lb-mole}}\right) \tag{5}$$

$$= 128\ \text{lb-mole/hr}\ (57.8 \times 10^3\ \text{g-mole/hr})\ SO_2$$

Sorbent System Design--

Several materials have been suggested for use in a dry sorbent baghouse flue gas desulfurization program. These include trona and nahcolite, which are naturally occurring sodium bicarbonate formations, pure sodium bicarbonate, and magnesium oxide. Of these, nahcolite has proven to be the most reactive, and therefore the most advantageous to use. Nahcolite has therefore been the sorbent material included in this design study. This is not meant to imply that nahcolite is the only economically feasible sorbent for this application, but rather that it is the only substance that has been studied and tested in sufficient depth to provide adequate data on material performance on which to base a full scale plant design.

Mechanism of Reaction Process--Dry process desulfurization with nahcolite involves the reaction of $NaHCO_3$ and Na_2CO_3 contained in the nahcolite ore with SO_2 in the flue gas stream. Genco and Rosenberg (1976) postulate that desulfurization occurs via the following 4 reactions:

$$2NaHCO_3(s) + SO_2(g) \rightleftarrows Na_2SO_3(s) + CO_2(g) + H_2O(g) \qquad (6)$$

$$2NaHCO_3(s) \rightleftarrows Na_2CO_3(s) + H_2O(g) + CO_2(g) \qquad (7)$$

$$NaCO_3(s) + SO_2(g) \longrightarrow Na_2SO_3(s) + CO_2(g) \qquad (8)$$

$$Na_2SO_3(s) + 1/2\ O_2(g) \longrightarrow Na_2SO_4(s) \qquad (9)$$

Reaction (7), the decomposition of $NaHCO_3$ to form Na_2CO_3, proceeds much faster than the reaction of SO_2 with $NaHCO_3$(6) or Na_2CO_3(8) (Bechtel 1976). Genco and Rosenberg (1976), simulating typical dry process desulfurization conditions, showed that in spent nahcolite a 1:1 ratio of Na_2CO_3 to Na_2SO_3 and a 2:1 ratio of Na_2SO_3 to Na_2SO_4 exists. Based on these data, the expected ratios of $NaHCO_3$, Na_2CO_3, Na_2SO_3, and Na_2SO_4 are, respectively, 1:12:26:10 in spent nahcolite.

Sizing of Injected Sorbent--It is apparent that the reaction of SO_2 with nahcolite is affected by several factors. First is the speed with which the $NaHCO_3$ is decomposed into Na_2CO_3. This process is clearly a direct function of particle size, with the finer materials decomposing at a significantly faster rate. Figure 4, which illustrates this observation, has been developed from data taken by the Superior Oil Company tests with a countercurrent reactor at the Cherokee Station of Public Service of Colorado(8). It can be seen from these data that for a fixed bed of sorbent, such as a collected layer in a baghouse, the rate of reaction of SO_2 with nahcolite may become limited by the rate of decomposition of sodium bicarbonate to sodium carbonate. To prevent this situation, it is necessary to maintain the sorbent size below a certain maximum particle diameter, which is determined by physical and chemical constraints of the operating system. Since standard baghouse technology will be utilized for the dry sorbent baghouse FGD system, it is necessary to provide adequate flexibility for variations in baghouse cleaning cycle timing. This cycle timing is determined by pressure drop across the baghouse

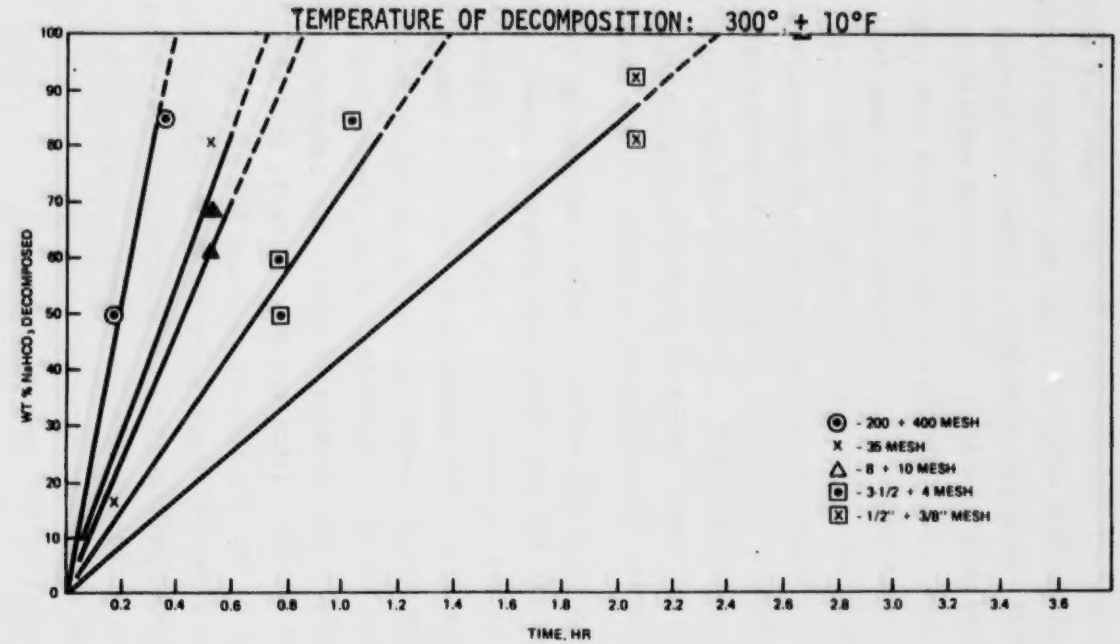


FIGURE 4. KINETICS OF NAHCOLITE DECOMPOSITION TO Na_2CO_3 (8). (FIXED BED TEST DATA TAKEN DURING NOV.-DEC., 1973). THE SUPERIOR OIL COMPANY.

compartments and may be required at intervals as frequent as 40 minutes. Based on the Superior tests at Cherokee, it is necessary to provide the nahcolite at approximately -200 mesh to ensure a supply of reactive material for reaction with SO_2 at 300°F.

Superior's test at Cherokee also included some test work on combustion zone injection of sorbent material. One very important finding of this test procedure was that nahcolite underwent a process of thermal comminution during its combustion zone residence. This process results in a cracking of nahcolite particles, thus providing a considerable size reduction. Figure 5 illustrates this phenomenon. It is interesting to note that the mass mean diameter of sorbent particles was reduced by a factor of 38, from 68 microns to 1.8 microns. Although it is possible to take advantage of this phenomenon by reducing the amount of sorbent preparation necessary before injection, it is doubtful that many utilities would agree to sorbent injection within the boiler due to the increased probability of slagging problems as a result of a significant increase in noncombustible solids in the combustion area.

Temperature of Reaction--Various tests performed on the dry sorbent process have demonstrated the strong dependence of removal efficiency on the gas temperature at the reaction site. Figure 6 illustrates this relationship based on test data taken with nahcolite by the Air Preheater Company during their tests at the Mercer Generating Station of Public Service Electric and Gas Company of New Jersey. Increasing the gas temperature dramatically increases the reaction rate, especially within the temperature range of 250° to 600°F (120° to 320°C). Since economic criteria will ultimately determine the applicability of the dry sorbent baghouse FGD process, it is necessary to consider here the effect of temperature on operating cost. To maintain a given required removal efficiency, the stoichiometric ratio can be reduced if the temperature is increased by a corresponding amount. This effectively reduces the amount of sorbent material required per ton of coal combusted, thus directly decreasing the operating costs of:

1) Nahcolite supply
2) Waste disposal

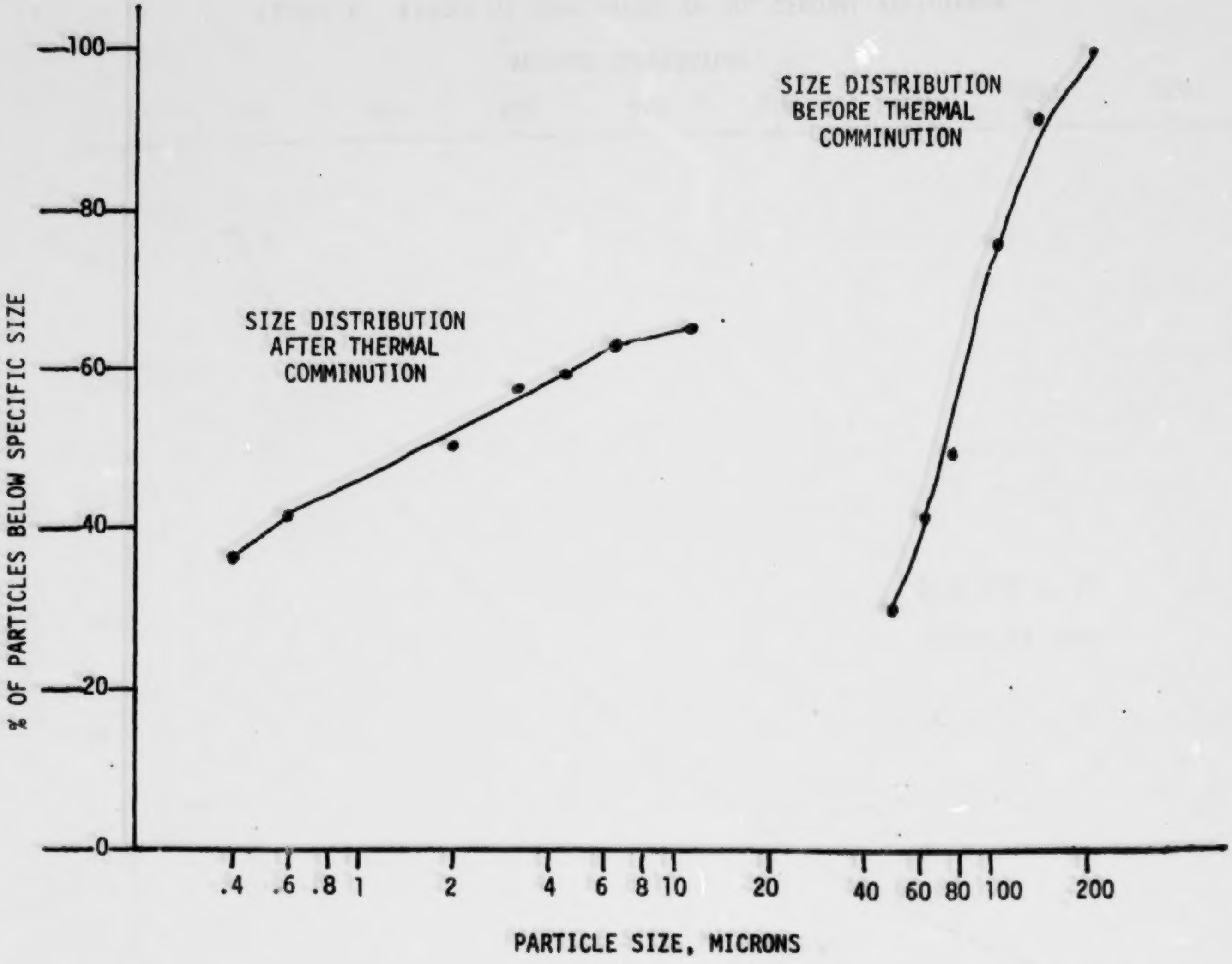


FIGURE 5. EFFECT OF THERMAL COMMINUTION ON PARTICLE SIZE OF SORBENT MATERIAL

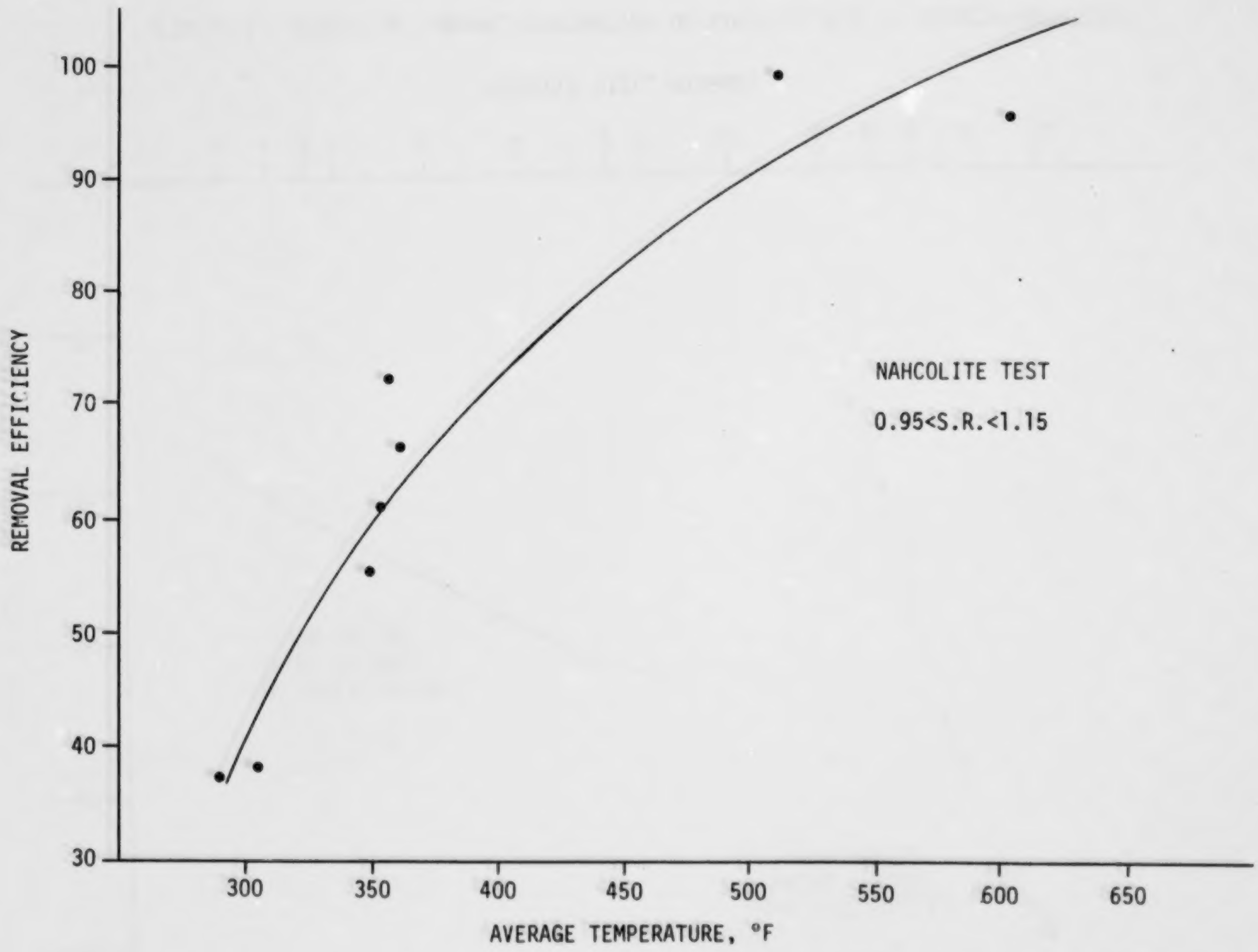


FIGURE 6. EFFECT OF TEMPERATURE ON SO_2 REMOVAL EFFICIENCY

Its effect on capital costs will be mixed, decreasing the cost of solids handling and waste disposal facilities and increasing the costs of the baghouse and air preheater sections.

The upper limit on reaction temperature is principally due to limitations on baghouse design. Standard baghouse designs and bag materials are generally applicable up to temperatures in the neighborhood of 550°F (288°C). To be consistent with current design philosophy, 400°F (204°C) has been chosen as a base design parameter; however, it is realized that operation at this temperature will provide less than optimum operation. This is due to a decrease in overall plant thermal efficiency dictated by the discharge of vast quantities of waste heat. The economic advantage of operation at an elevated temperature in the neighborhood of 550°F (288°C) is illustrated in a subsequent section of this report dealing directly with flue gas desulfurization economics.

Stoichiometric Ratio--The required removal efficiency of the FGD system is a function of the chemical analysis of the as-fired coal, the boiler efficiency, and the applicable Federal and state regulations. Based on the assumptions stated earlier, a 70 percent SO_2 removal efficiency is required for this design.

Removal efficiency and reaction temperature of the dry sorbent system determine the average stoichiometric ratio for the sorbent system. Contact time will also exert an influence on the stoichiometric ratio, but this effect is relatively small when compared to efficiency and temperature within the range of practical variations in contact time. Figure 7 illustrates this relationship based on test data by The Air Preheater Company at Mercer Generating Station(9) and by American Air Filter Company(10) during bench-sale test work. From these experimental results, it can be seen that at 400°F (204°C), a SO_2 removal efficiency of 70 percent will require a stoichiometric ratio of approximately 1.0. At a stoichiometric ratio of 1.0, the average utilization of nahcolite will equal the efficiency of 70 percent. This factor identifies the quantity of sorbent necessary by determining the net loss of unreacted sorbent as being 30 percent.

It should be noted here that the determination of stoichiometric ratio has been based on the experimental data from a limited number of tests. Data not published at the time of this report suggest that higher efficiencies

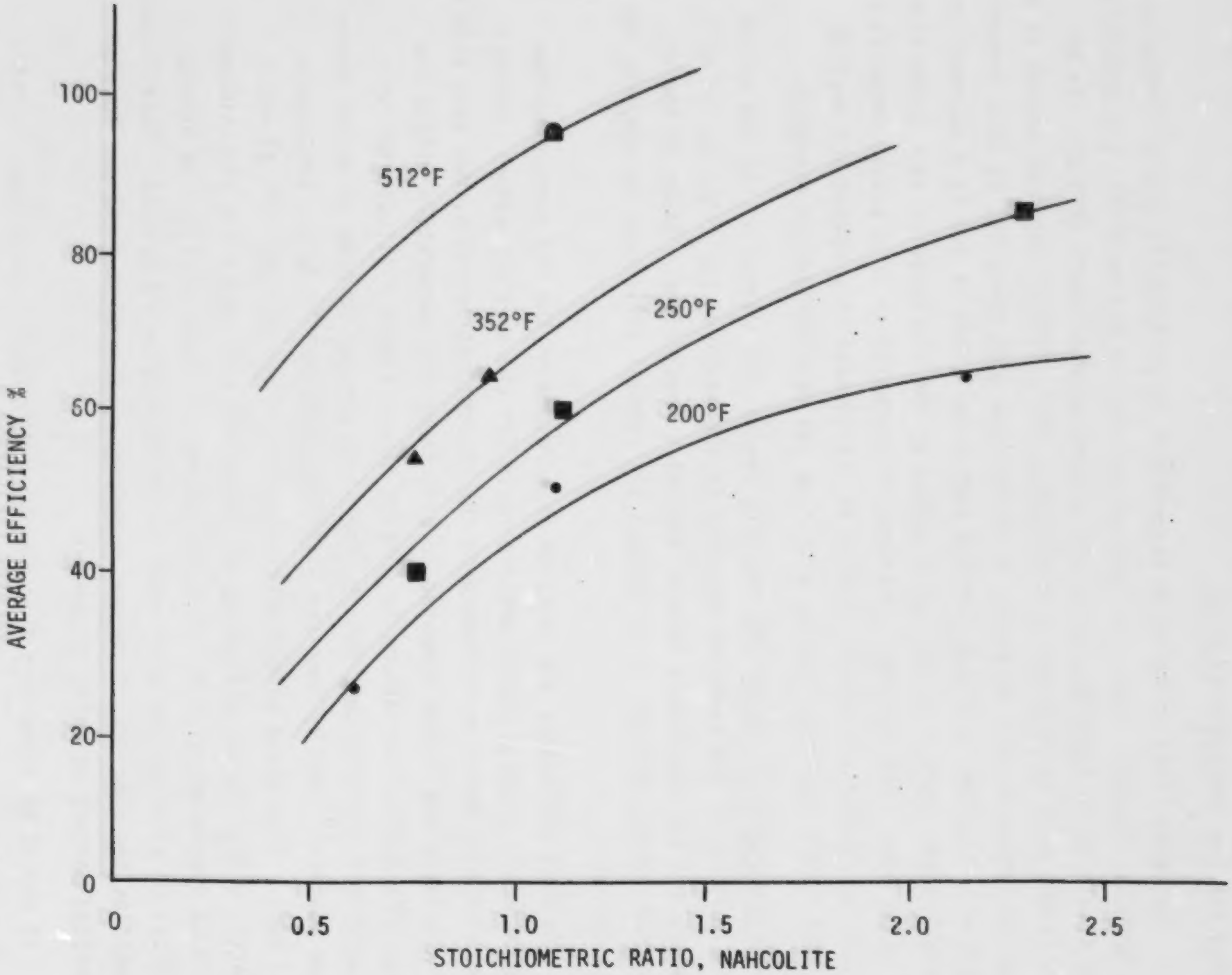


FIGURE 7. EFFECT OF STOICHIOMETRIC RATIO ON REMOVAL EFFICIENCY

may be attainable for a given Stoichiometric Ratio and temperature than suggested by Figure 7. Results of this type indicate that a Stoichiometric Ratio below 1.0 may be used for the dry sorbent baghouse process. This will increase the sorbent utilization, thus decreasing the amount of sorbent required and hence reduce the FGD system operating costs.

Contact Time--Air Preheater Company(9) found "There is a definite advantage by injecting the additive further upstream of the filter house to increase retention time." Their study found increases of from 17 to 27 percent in SO_2 removal efficiency. To prevent possible equipment damage due to the high particle load of the flue gas stream containing both fly ash and nahcolite, the injection point for the sorbent should not be upstream of any air preheater or other flue gas handling equipment. The principal injection point for the base design is therefore immediately downstream of the air preheater.

Sorbent Usage--As previously determined, the base design plant will have a flue gas concentration of 128 moles/hr of SO_2. The amount of nahcolite required for plant operation is therefore determined as that amount required to react 70 percent (the required removal rate) of this flue gas component. The amount of SO_2 removal is:

$$Q_{mole/hr} = (128 \text{ mole/hr})\ (0.70) \qquad (10)$$

$$= 88 \text{ mole/hr}$$

Two moles of $NaHCO_3$ are required to react chemically with one mole of SO_2. The quantity of $NaHCO_3$ required to react with 88 moles/hr of SO_2 is:

$$Q_{mole/hr} = (88 \text{ mole/hr})\ (2) \qquad (11)$$

$$= 176 \text{ mole/hr } NaHCO_3$$

At a stoichiometric ratio of 1.0, the nahcolite utilization equals the SO_2 removal efficiency of 70 percent. The amount of $NaHCO_3$ required is therefore increased to allow for this utilization:

$$Q_{mole/hr} = (176 \text{ mole/hr}) \left(\frac{1}{0.70}\right) \qquad (12)$$

$$= 250 \text{ mole/hr NaHCO}_3$$

One mole of $NaHCO_3$ weighs 84 pounds (38 kg). Therefore, 250 mole/hr of $NaHCO_3$ is equivalent to:

$$Q_{lb/hr} = (250 \text{ mole/hr}) (84 \text{ lb/mole}) \qquad (13)$$

$$= 21{,}000 \text{ lb/hr } (9{,}500 \text{ kg/hr})$$

Nahcolite ore is assumed to assay at approximately 70 percent $NaHCO_3$. The quantity of nahcolite required for this process is therefore given by:

$$Q_{lb/hr} = (21{,}000 \text{ lb/hr}) \left(\frac{1}{0.70}\right) \qquad (14)$$

$$= 30{,}000 \text{ lb/hr nahcolite } (13{,}600 \text{ kg/hr})$$

Injection Method--The principal injection point for the sorbent is immediately downstream of the air preheater. Tests performed on this configuration with continuous injection identify an initial period after the commencement of the test during which the removal efficiency is considerably reduced. A typical test result of this type is presented in Figure 8. This phenomenon can be explained by the observation that the actual SO_2 removal takes place both in the flue gas stream and on the bag surface. Removal efficiency will therefore be at a reduced level until the nahcolite coating on the bag reaches some minimum coating thickness. Optimum performance of the dry sorbent FGD system must therefore require a combination of nahcolite injection immediately downstream of the air preheater and some preliminary precoating of the filter bags. This configuration will prevent the initial removal efficiency loss and provide the most beneficial removal efficiency during continuous operation.

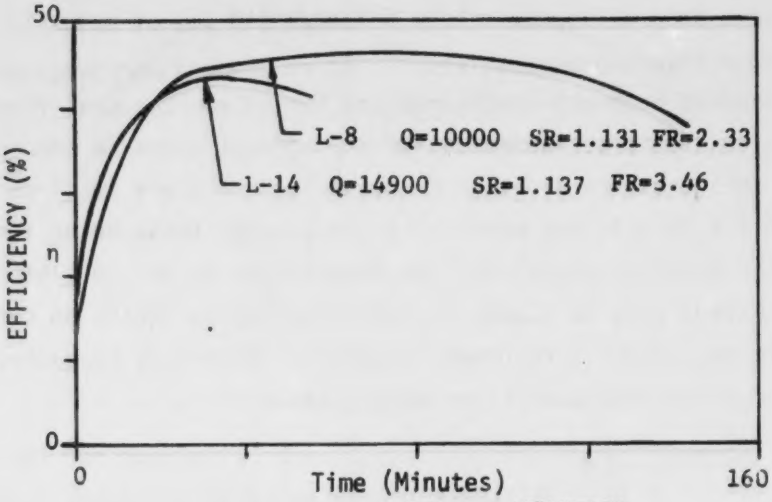


FIGURE 8. REMOVAL EFFICIENCY(9)

L = Test Designation
Q = Flow Rate, acfm
SR = Stoichiometric Ratio
FR = Filter Ratio, Ft^3M/Ft^2

Air to Cloth Ratio--The air to cloth ratio (A/C) is defined as the actual flow rate of air passed through the baghouse in cubic feet per minute divided by the active cloth filter area in square feet. The active cloth area is the actual bag surface passing flue gas at any instant, and does not include unused surfaces such as bags undergoing cleaning. Air to cloth ratios will affect deposit characteristics, collector efficiency, pressure drop, and maintenance requirements. Of these, the most important is pressure drop requirements. Greater A/C ratios will increase the differential pressure across the bag surface. Induced draft fans are required to move the flue gas through the baghouse. The capital costs of these fans increase with pressure drop, as do the operating costs. Dry sorbent baghouses will be located upstream of the induced draft fans, and will therefore be operated at less than atmospheric pressure. The magnitude of this negative pressure is directly proportional to the differential pressure across the filter bags. Construction costs for the baghouse structure increase rapidly with pressure drop due to the increase in structural rigidity necessary to withstand the differential pressure without structural damage.

A pressure drop of approximately 4 inches (10 cm) of water is considered by Wheelabrator-Frye representatives(21) to be the optimum pressure drop attainable with standard baghouse configurations for pulverized coal-fired power plants. Various tests performed on the dry sorbent baghouse concept demonstrate that this pressure drop can be maintained with air to cloth ratios ranging from 2.4 to 3.4; the main variation between these being the cycle timing for the baghouse operation. We have chosen an air to cloth ratio of 3.0 for the base design to allow reasonable operating limits on cycle timing. This value is consistent with recent commercial proposals by baghouse manufacturers for pulverized coal-fired power plants.

Precoat Sorbent Usage--The quantity of nahcolite required for use in precoating the filter bags is fundamentally based on achieving a minimum precoat thickness. Tests by Air Preheater Company(9) on a dry sorbent baghouse with all bags cycled in parallel demonstrated a specific removal efficiency time dependence as illustrated in Figure 9. The active filter bag area was 4,300 ft^2 (400 m^2) and the bag was coated with sorbent at the rate of 25 lb/min (11 kg/min). From Figure 9, peak efficiency was reached after 1.7 minutes had elapsed. The amount of sorbent material that had deposited on the filter bags at this time can be calculated as:

$$A_{lb/ft^2} = (1.7 \text{ min})\ (25 \text{ lb/min})\ \left(\frac{1}{4{,}300 \text{ ft}^2}\right) \qquad (15)$$

$$= 0.01 \text{ lb/ft}^2\ (0.05 \text{ kg/m}^2)$$

This is the amount of precoat which is therefore necessary to minimize the effect of initial SO_2 removal loss.

At an air to cloth ratio of 3.0, the base design flow of 1.5×10^6 acfm (4.2×10^4 m^3/min) will require 5.0×10^5 ft^2 (4.6×10^4 m^2) of active bag area. Assuming a baghouse cycle time permitting an average on-stream bag exposure of 50 minutes per cycle, the nahcolite required for the filter bag precoating is given as follows:

$$Q_{lb/hr} = (0.01 \text{ lb/ft}^2)\ (5.0 \times 10^5 \text{ ft}^2/50 \text{ min})\ (60 \text{ min/hr}) \qquad (16)$$

$$= 6{,}000 \text{ lb/hr}\ (2{,}724 \text{ kg/hr})$$

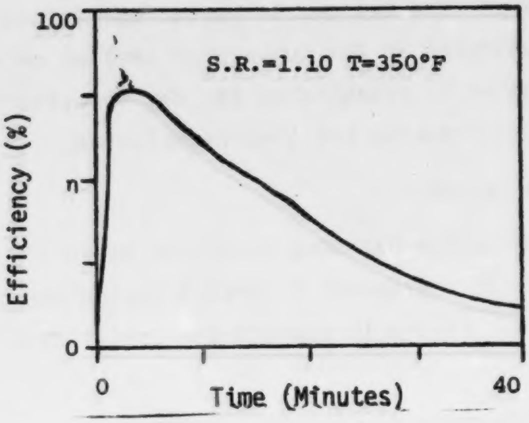


FIGURE 9. REMOVAL EFFICIENCY VS. TIME(9)

Injection Sorbent Usage--The amount of nahcolite required for injection downstream of the air preheater is therefore the balance of the material, or 24,000 lb/hr (10,876 kg/hr).

Total Material Collected by Baghouse--Specifications for several of the baghouse design parameters have been developed in the preceeding sections. An additional parameter required for the baghouse design is the total amount of material that will be collected on the filter bags. This total will be a combination of the fly ash and dry sorbent. The sorbent consumption rate will be 30,000 lb/hr (13,600 kg/hr) and the fly ash production rate will be 36,550 lb/hr (16,575 kg/hr). The resultant solids loading on the filter bags will be 66,550 lb/hr (30,175 kg/hr), or 0.097 lb/ft^2 (0.47 kg/m^2) of bag surface.

Physical Design

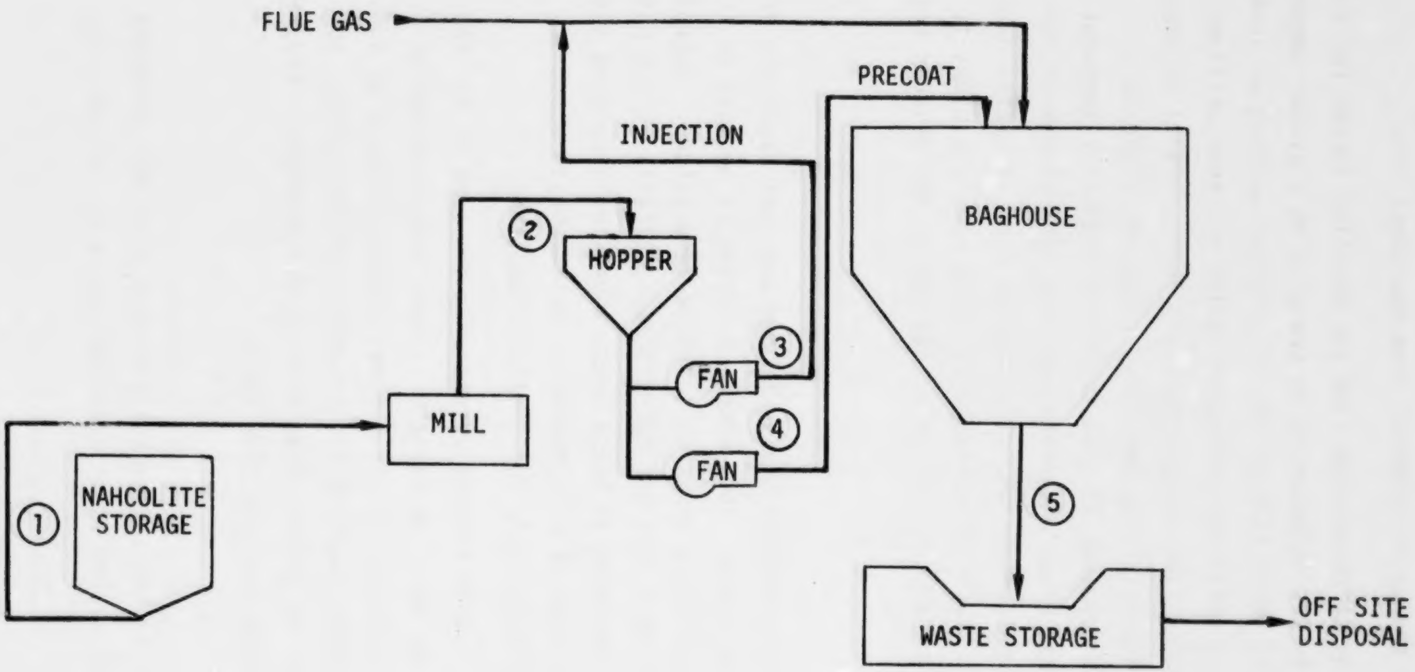
A physical design for the dry sorbent baghouse flue gas desulfurization system must be established to ascertain equipment needs and develop an economic model. Design methods and sizing criteria follow standard procedures for the power industry; however, in process storage capacity and equipment, spares have been minimized for consistency with the design philosophy present in the TVA study, "Detailed Cost Estimates for Advanced Effluent Desulfurization Processes." The design data presented in the following sections have been used only to establish a base design system. The sensitivity of the capital and operating costs of this design, to variations in process design and coal parameters, is illustrated in the subsequent section on economics. Once an optimum operating point is established for any specific installation, the specific hardware requirements are then reevaluated.

Layout, Base Design System--

The base design system has been developed as an illustrative tool for the dry sorbent system. It represents a careful evaluation of all existing test work on this process. Figure 10 depicts the solids handling portion of the plant.

Nahcolite is delivered to the plant by railroad in covered hopper cars. Cars are unloaded semiautomatically and the nahcolite is transferred to covered silos for normal plant storage. The railroad unloading facility is sized to accommodate 75 tons (68 metric tons) per hour, based on operating on a 40 hour work week. All nahcolite transfer/conveying systems, both before and after the milling operation, are of the pneumatic conveying type. The nahcolite transfer system from railroad unloading to site storage will include an automatic switching system to distribute the sorbent among four 200,000 cu. ft. (5,663 cu. meter) storage silos. This provides an on-site sorbent storage of 32 days plus whatever amount is contained within the railroad hopper cars not yet unloaded, typically 3 to 5 day's worth. Nahcolite is transferred from these silos to a surge tank which feeds directly into the mill. The surge tank is sized for a capacity of 8,000 cu. ft. (230 cu. meter), or eight hours of operation. Two mills are provided, each having a capacity of milling 20,000 lb/hr (9,000 kg/hr) of nahcolite to -200 mesh (less than 0.074 mm). Included with the mill is a 4,500 cu. ft. (130 cu. meter) receiver tank which collects the milled sorbent.

Two rotary feeders draw the milled sorbent from the receiver tank and meter it to the two points of nahcolite injection. Sorbent and fly ash are collected on filter bags within the baghouse and are subsequently convenyed to a waste material silo for removal. The baghouse is designed to handle 1.5 x 10^6 acfm (4.25 x 10^4 cu. meter/min) of flue gas at 400°F (200°C) at an air to cloth ratio of 3:1. The waste material silo is sized to accommodate 300,000 cu. ft. (8,500 cu. meter) of combined sorbent/fly ash material, which



	1 CRUSHED NAHCOLITE	2 MILL OUTLET	3 NAHCOLITE TO INJECTION	4 NAHCOLITE TO PRECOAT	5 SPENT ABSORBENT/ FLY ASH
FLOW (lb/hr)	30,000	30,000	24,000	6,000	64,400

FIGURE 10. FLOW CHART - SOLIDS HANDLING SYSTEM, BASE DESIGN

will allow an accumulation of 5 days' worth of operation without disposal. Trucks will be used to haul the waste material from the plant site.

Figure 11 represents the corresponding flue gas handling system for the base design system. To allow the baghouse to be operated at elevated temperatures in the range of 400 to 550°F (200 to 290°C), without wasting an excessive amount of heat energy, a split air preheater design has been utilized. Flue gas exiting from the economizer at approximately 705°F (375°C) is cooled to the baghouse operating temperature by the first stage air preheater. Cleaned flue gas exiting the baghouse is then cooled to the stack temperature of 225°F (110°C) by the second stage air preheater, the heat recovered from both stages being used to preheat the boiler combustion air. A bypass damper is provided around the first stage air preheater so that when the plant is operated at reduced loads the baghouse can be maintained at the desired temperature.

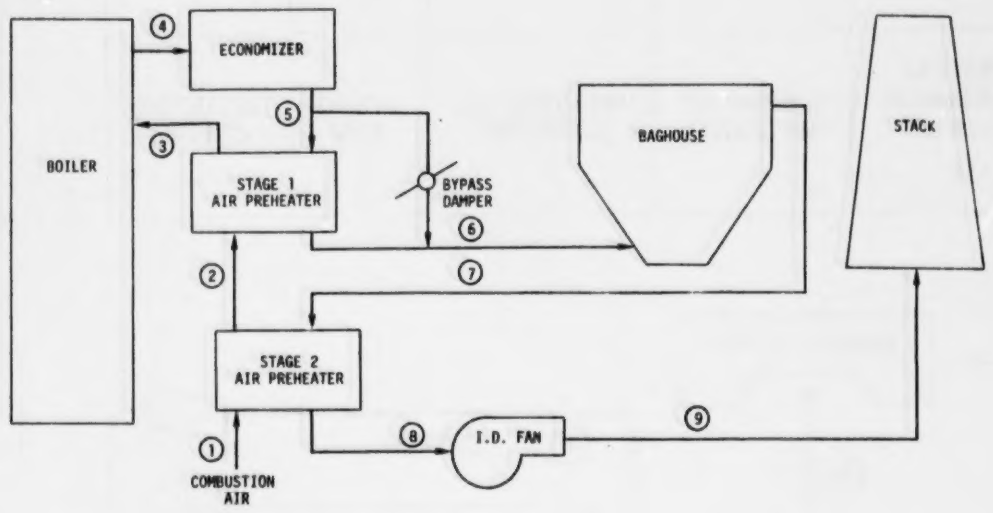
Layout, Retrofit System--

The economics of the dry sorbent baghouse system are most favorable for design inclusion within a new plant. This occurs principally because of the sensitivity of capital and operating costs to baghouse temperature. Retrofit systems must operate at available flue gas conditions, typically in the neighborhood of 275°F (135°C). Operation at these temperatures will be more costly and the dry sorbent/baghouse concept will therefore maintain its economic advantage over other FGD processes only for limited situations.

The nahcolite handling system remains essentially the same as for the base design system except that the flow rates will vary from that design as illustrated in Figure 12. Injection of dry sorbent material should be downstream of any existing equipment, such as electrostatic precipitators, to prevent any solids collection at points other than in the baghouse. Figure 13 represents the corresponding flue gas flow chart.

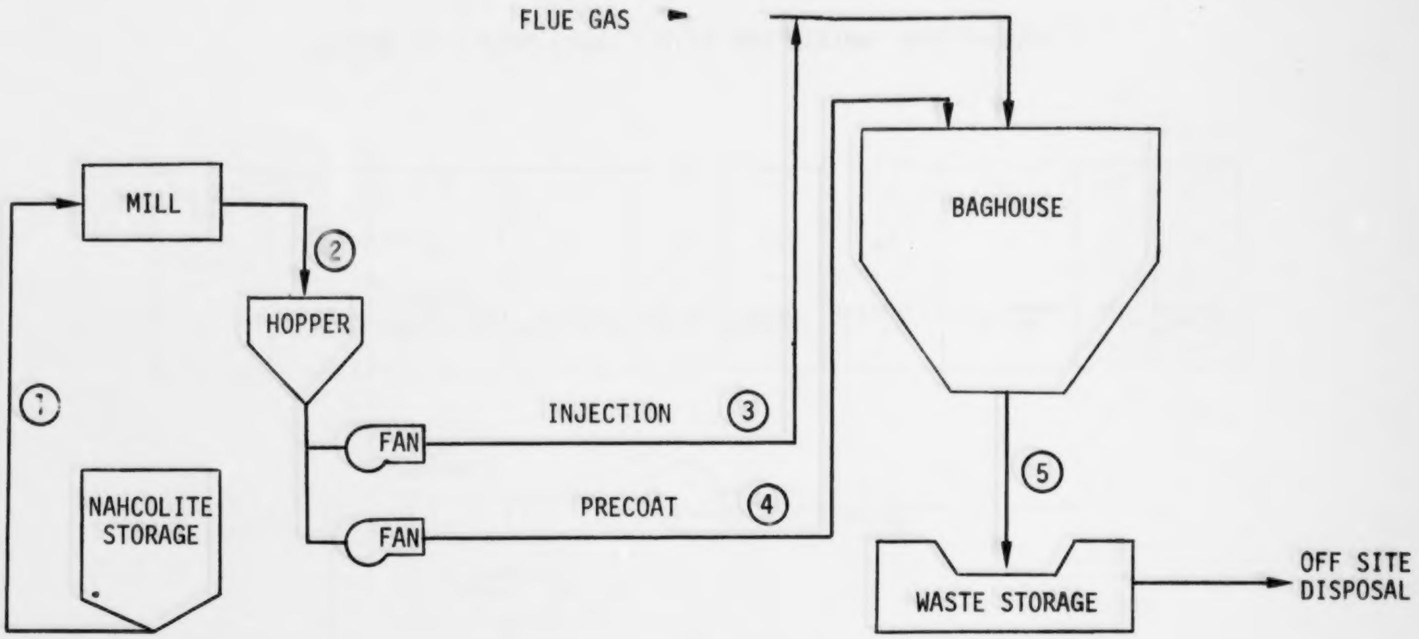
SORBENT SUPPLY AND COSTS

Of paramount importance in the successful utilization of the proposed dry sorbent-fabric filter FGD system is an adequate supply of suitable dry



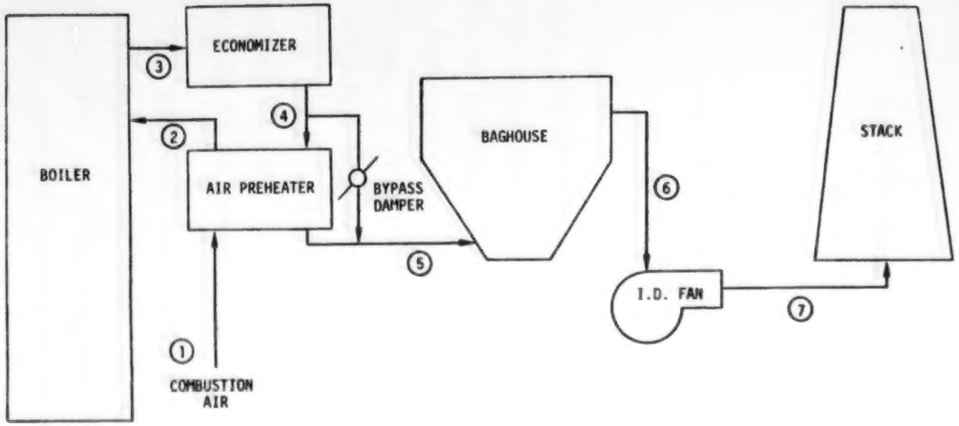
	1 COMBUSTION AIR INLET	2 COMBUSTION AIR TO STAGE 1 AIR PREHEATER	3 COMBUSTION AIR TO BOILER	4 BOILER FLUE GAS	5 ECONOMIZER OUTLET	6 FLUE GAS TO BAGHOUSE	7 FLUE GAS FROM BAGHOUSE	8 I.D. FAN INLET	9 FLUE GAS TO STACK
TEMPERATURE (°F)	110	292	655	890	705	400	375	225	225
FLOW (10^6 lb/hr)	4.0	4.0	4.0	4.3	4.5	4.7	4.7	4.9	4.9

FIGURE 11. FLOW CHART - FLUE GAS SYSTEM, BASE DESIGN



	1 CRUSHED NAHCOLITE	2 MILL OUTLET	3 NAHCOLITE TO INJECTION	4 NAHCOLITE TO PRECOAT	5 SPENT ABSORBENT/ FLY ASH
FLOW (lb/hr)	65,000	65,000	59,470	5,530	99,400

FIGURE 12. FLOW CHART - SOLIDS HANDLING SYSTEM, RETROFIT DESIGN



	1 COMBUSTION AIR INLET	2 COMBUSTION AIR TO BOILER	3 BOILER FLUE GAS	4 ECONOMIZER OUTLET	5 FLUE GAS TO BAGHOUSE	6 FLUE GAS FROM BAGHOUSE	7 FLUE GAS TO STACK
TEMPERATURE (°F)	110	550	890	705	275	263	260
FLOW (10^6 lb/hr)	3.8	3.8	4.1	4.4	4.7	4.7	4.7

FIGURE 13. FLOW CHART - FLUE GAS SYSTEM, RETROFIT DESIGN

sorbent at a reasonable cost. This section addresses both the cost and availability of the most promising of the sorbents, nahcolite.

Scope of Analysis

Some materials such as raw trona (sodium sesquicarbonate [$Na_2CO_3 \cdot NaHCO_3 \cdot 2H_2O$]) could have potential for use as a dry SO_2 sorbent, however, these other sorbents are not considered in the present analysis for one or more of the following reasons:

- Insufficient reactivity and other performance-related data were available for process economics to be assessed in some cases, such as trona.
- Other potential sorbents, such as magnesium oxide, limestone, and lime, were eliminated because of their low reactivities and the consequent high sorbent consumption and waste generation rates.

It should be noted that trona has at least one significant advantage over nahcolite: it is currently being mined in the Green River Basin of Wyoming in large quantities by such companies as Allied Chemical Corporation, FMC Corporation, Stauffer Chemical Corporation, and Texasgulf, Incorporated. Although the trona is used only as a raw material for soda ash (Na_2CO_3) manufacture at this time, it is not difficult to envision the sale of raw trona for FGD purposes should a sufficient market develop.* In view of trona's ready availability, an experimental program for the baghouse FGD concept should include at least brief tests of trona to define its performance as an SO_x sorbent.

The western United States includes several known deposits of nahcolite, but for FGD purposes, the economically significant nahcolite deposits are all found in the Piceance Creek Basin of northwestern Colorado (Figure 14). The deposits are part of the Green River Formation, which includes the Green River Basin of southwestern Wyoming.

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*However, it is difficult to make estimates of the possible market price of raw trona, because the present depletion allowance is based in part on the value of the final product, soda ash.

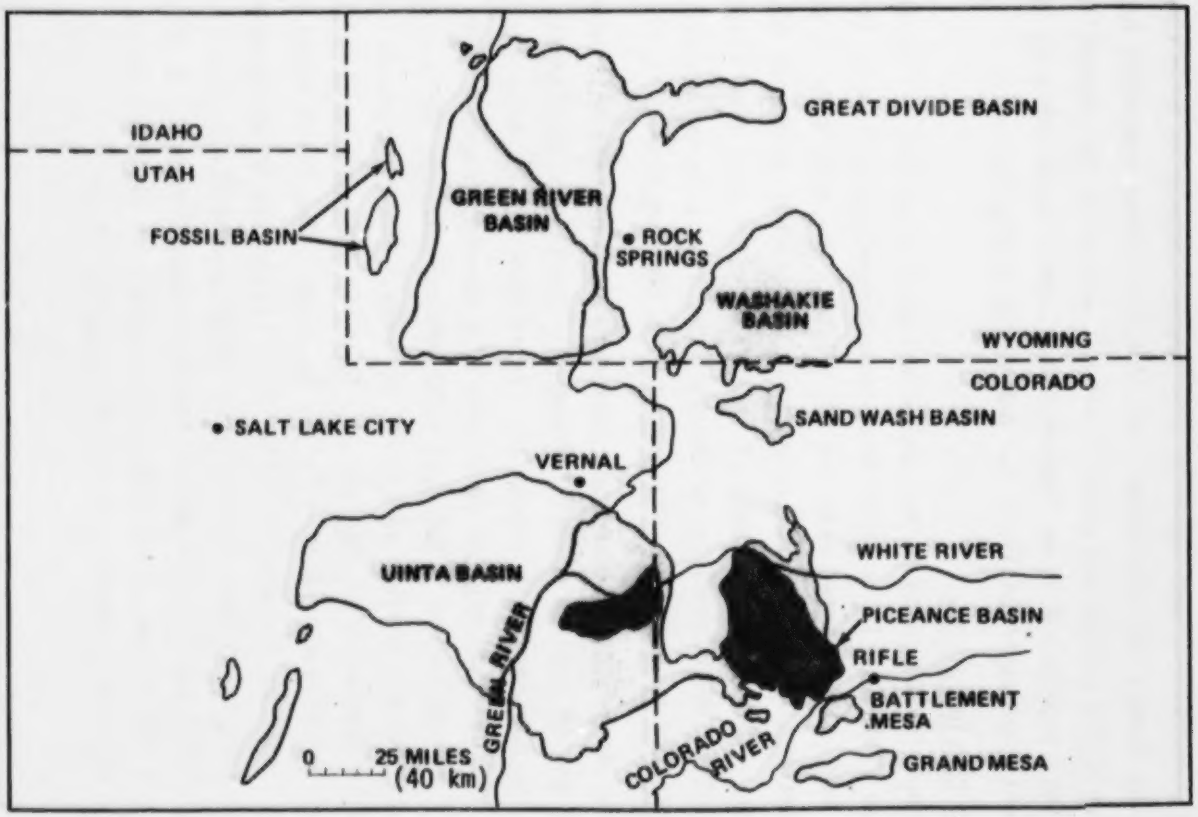


FIGURE 14. GREEN RIVER FORMATION (DARK SECTION IDENTIFIES THE MAHOGANY ZONES)(8)

This geographic area has therefore been chosen to base the present analysis on.

There are numerous sodium lease holders in the Piceance Creek region; however, three companies--Industrial Resources, Incorporated (IRI), Rock School Corporation, and Superior Oil Company--appear to hold some of the most promising leases for nahcolite. In addition, these companies have all prepared detailed mining and production cost estimates for nahcolite. These estimates, together with an independent economic assessment by TRW, constitute the overall economic analysis for nahcolite production.

General Considerations

Nahcolite is found in the unleached portion of the oil shales of the Colorado Plateau. It is found both in a disseminated form in the oil shale and in beds interlayered with the oil shale. The beds vary in purity from 30 to 90 percent nahcolite and range from a few inches to twenty plus feet in thickness. Overlying the beds of interest, which vary from 1,500 to 2,500 feet (460 to 760 meters) below the surface, lies a large aquifer which must be penetrated in order to obtain access to the nahcolite.

The construction techniques necessary to penetrate this highly permeable aquifer and to operate a shaft through it constitute the largest unknown factor in the mining of the nahcolite. In recognition of this fact, the U.S. Bureau of Mines has drilled an 8-foot (2.4 meters) diameter test shaft to a depth of approximately 2,300 feet (700 meters) through the aquifer to the bottom of the oil bearing shales in the area. The experimental shaft is located in Horshoe Draw, Section 30, Township 1 South, Range 97 West about 30 miles (48 kilometers) from Rifle, Colorado. The first shaft station will be operational by the end of 1978. This station is in the embedded layer of nahcolite, 1,660 feet (500 meters) below ground level, and is expected to yield 1,100 tons (1,000 metric tons) of nahcolite ore. A second shaft station in the disseminated layer of nahcolite is expected to be operational early in 1979, producing a similar quantity of ore.

The second largest cost factor that has to be considered as differentiating oil shale mining from some other mining modes is the matter of possible explosive atmospheres. The Mining Enforcement and Safety Administration (MESA) had tentatively assigned a "gassy" mine classification to oil shale mining, a classification which automatically specifies certain operational requirements for egress and ventilation, as well as the use of equipment which has passed Bureau of Mines "permissibility" (ability to operate in an explosive atmosphere) testing schedules. During the 10th Oil Shale Symposium at the Colorado School of Mines in April 1977, a paper was presented on the expected dust/fume atmosphere of an oil shale mine. If these expectations hold true, it will mean an additional 25 to 50 percent cost for underground equipment.

Nahcolite Reserves and Mining Methods

As previously noted, the Piceance Creek Basin of the Green River Formation in northwestern Colorado contains the principal nahcolite deposits as shown in Figure 14. The mahogany zone is a kerogen-rich oil shale layer which overlies much of the richer nahcolite deposits. Because of the mutual occurrence of nahcolite with oil shale and dawsonite ($NaAlCO_3(OH)_2$), some lease holders have postulated the joint mining of all three materials. This would be of obvious benefit in developing nahcolite production.

Nahcolite Reserves--

Oil shale exploration in the Piceance Creek Basin has resulted in the production of numerous core drillings. Based on the assays of these core samples, the Bureau of Mines estimates the total nahcolite reserves in the basin to be about 30 billion tons (27 billion metric tons)(8). Much of this is either unrecoverable or of very low assay; however, the total reserves appear to be easily adequate for FGD applications.

Several organizations have developed mining plans for nahcolite, and the potential supply should satisfy even extensive application of the dry sorbent-baghouse system. Maximum production rates from some of the prominent companies are shown in Table 4.

TABLE 4. PLANNED MAXIMUM NAHCOLITE PRODUCTION RATES [8,22,23]

Organization	Production Rate Tons/Year (Metric Tons/Year)	Mine Life Years
Cleveland Cliffs Iron Co. (Rock School Corp.)	481,800 (437,080)	15
Industrial Resources, Inc.	2,273,000 (2,062,000)	40*
Superior Oil Co.	1,095,000 (993,400)	~25†

*This estimate is based on the recoverable reserves in the nahcolite bed to be developed.

†This is for nahcolite-only case without coproduction of oil shale.

The total production from the above three mines alone is sufficient to satisfy the nahcolite demand of over thirty-five 500 MW coal burning plants with the characteristies of the TRW base case design. If the demand develops, there seems little doubt that the nahcolite supply will also be developed. It should be noted that the nahcolite from the IRI mine is of high assay (>70%) and can be sold on a run-of-mine (ROM) basis. The other two mines produce lower assay material which requires concentration prior to sale. The production rates given are for the as-sold nahcolite.

Mining Methods--

The conventional mining schemes proposed for use in mining the bedded nahcolite parallel the standard room and pillar methods used for mining other bedded deposits such as salt, potash, trona and coal. Structurally, the pillar requirements will more closely parallel the mining of potash and trona which are mined at similar depths. Because of the MESA assignment of a fire hazardous situation for mining in oil shale, the entry, exit and ventilation requirements would more closely parallel coal mining.

A standard mining cycle would be followed. It consists of drilling, placing explosives and blasting; loading the ore; transporting the ore; placing roof support; and repeating the cycle. Each of these portions of the cycle require the exclusive use of a room to avoid interference and keep the various crews continuously occupied. In order to keep a continuous flow of ore to the secondary and main haulage systems, a large number of these working sections are required. At maximum production, the cycle would be expected to produce the equivalent of two rounds (300 tons(270 metric tons)) per shift or 600 tons (540 metric tons) per section per day. Thus, with four rooms to a section, a mine producing 10,000 tons (9,000 metric tons) per day in two shifts (245 working days per year or approximately 2.5 million tons [2.3 million metric tons] per year) would require some 17-18 working sections.

The operation of the basic mining cycle is best exemplified by a review of the plans of Industrial Resources, Inc. (IRI) as presented to TRW by the courtesy of their management. IRI plans to open a nahcolite mine in the area of Township 7 South, Range 101 West, in Colorado. They plan to use a two boom jumbo drilling rig with a rotary drill on each boom for producing forty 1-1/2 inch (3.8 cm) diameter holes for explosives detonated for each round to be used in breaking the ore.. After the ore is broken, it will be loaded and trammed to a feeder by a seven cubic yard (5.35 cubic meter) front end loader. From the feeder, the ore is transferred to a secondary haulage conveyor belt 36-inches (91 cm) in width which in turn feeds the 42-inch (107 cm) width main haulage belt. The main haulage belt feeds both a live storage area and intermediate storage silos (ore pockets underground) which in turn feed into ore pockets by the shaft which load into 25 ton (23 metric tons) skips for hoisting to the surface for crushing and/or further processing. To complete the cycle, after the broken ore is removed, IRI plans to install high-strength roof bolts on a 4 by 4.5 foot (1.2 by 1.4 meters) pattern in the rooms for support of the roof during subsequent mining operations.

The Cleveland Cliffs plan also calls for a conventional vertical mine, but the shafts are smaller (12-foot (3.6 meter) production shaft and a 7-foot (2.1 meter) access shaft) than those for the IRI plan. The Superior plan is significantly different in that inclined access will be used.

In the northeastern part of the Piceance Creek Basin, minable nahcolite strata appear quite near the surface at the North Outcrop. Superior plans to access the nahcolite by means of an 11,500 foot (3,505 meters) inclined tunnel This plan should substantially reduce the difficulties involved in sinking a vertical shaft through a fractured, highly permeable aquifer.

Reference was made earlier to the similarity of various mining methods. This analogy can be carried even further in the case of potash mining in New Mexico and Saskatchewan where the deposits were also overlain by large aquifers. In New Mexico, a method of shaft sinking for the potash mines was used which might prove applicable to the nahcolite deposits. In sinking the shafts to the potash, a series of holes were drilled and cased in a circle outside the planned dimensions of the shaft. Ammonia refrigerant was then circulated until the entire perimeter and much of the central area was frozen (two to three years) and the shaft was excavated through the frozen zone, lined and sealed and then the outer area was allowed to thaw. Most of these shafts were constructed many years ago and today it may be that chemical grouts could provide a simpler solution though the more recent Canadian mines also used freezing. In any case, this example points up the difficulties of this type of shaft sinking.

Mining Capital Costs and Nahcolite Pricing

Since there is no established market for nahcolite, its selling price can only be established by preparing a complete, detailed mine design and estimating the costs, or by analogy with similar types of mining and their costs. Because of the inherent limits of accuracy caused by the shaft sinking situation in the oil shale area, TRW selected to develop its cost estimates by analogy. On the other hand, Cleveland-Cliffs Iron Co., Industrial Resources, Inc., and Superior Oil Company all used detailed mine plans in developing their cost estimates. Consequently, the TRW estimates should be viewed as rough checks on the validity of the other cost analyses. The various cost estimates are summarized in Table 5.

Previous work by TRW established the costs of mining thick bedded coal by conventional room and pillar methods at $9.28 per ton ($10.23 per metric ton) without a shaft. Recent work by Bechtel for the U.S. Bureau of Mines

TABLE 5. MINE DEVELOPMENT COSTS AND NAHCOLITE PRICES

Organization	Production Rate* Tons/Yr	Production Rate* (Metric Ton/Yr)	Preproduction Costs	Sales Price, F.O.B. Mine $/Ton	Sales Price, F.O.B. Mine ($/Metric Ton)
Cleveland-Cliffs Iron Co.[†]	146,000	(132,000)	$ 8,727,000	25-30	(28-33)[∇]
	321,000	(291,000)[§]	9,316,000	20-30	(22-33)[∇]
	482,000	(437,000)[§]	9,316,000	20-30	(22-33)[∇]
Industrial Resources, Inc.[#]	500,000	(454,000)	$28,370,000	35	(39)
	1,000,000	(907,000)	35,770,000	25	(28)
	2,273,000	(2,062,000)	44,030,000	17	(19)
Superior Oil Co.**	1,095,000	(993,000)	$40,000,000	Not available	
TRW-Coal Analogy	2,500,000	(2,268,000)	$12,000,000	10.30	(11.35)
TRW-Manure Salts Analogy	2,500,000	(2,268,000)	12,000,000	13	(14)

*This is the rate for salable material, not necessarily the same as the gross mine output.

[†] These data were provided through the courtesy of Mr. J.H. Smith, Rock School Corp.(22)

[∇] The ranges result from differing assumptions regarding depletion allowances and return-on-investment.

[§] These production rates are based on the same physical plant, but the lower rate is a 2-shifts per day basis and the higher is a 3-shifts per day basis.

#These data were provided through the courtesy of Mr. Jacques Dulin, Industrial Resources, Inc.(23)

**These data were abstracted from reference (8).

estimated the costs of shaft sinking to the appropriate depth at approximately $4 million in stable ground with no problems. Factoring this amount to allow for unstable ground and adding the costs for the necessary headframe and hoists leads to a total of approximately $12 million. A discounted cash flow analysis with a 15 percent rate of return yields, over a 20-year mine life, approximately $1.00 per ton ($1.10 per metric ton) in additional add-ons to the selling price for a total price of $10.30 per ton ($11.35 per metric ton) F.O.B. the mine.

Potash is normally sold after processing for potassium chloride which is presently selling at $0.65 per unit ($K_2O$) or $39 per ton ($43 per metric ton). Manure salts (20% K_2O) is sometimes sold for a flux in the metals industry. It is a run-of-mine material with a minimal amount of processing (similar to crushed nahcolite) and sells for approximately a K_2O equivalency with the KCl or $13.00 per ton ($14.33 per metric ton) F.O.B. the mine.

Both of these industries are mature and highly competitive with underground coal mines competing with strip mining costs and the oversupply of potash from the Canadian mines driving that market.

Given these situations and the fact that a start-up industry such as nahcolite mining must command greater than a 15 percent discounted cash flow rate of return in order to attract capital, it would appear that the IRI estimates for a 2.3 million ton per year (2.1 million metric ton per year) mine of $17.00 per ton ($18.75 per metric ton) are reasonable. This is in good agreement with the Cleveland Cliffs sales price estimates ($20-30 per ton ($22-33 per metric ton)) when one considers the lower production rates of the latter operation and the higher assay product (∿80%). As the nahcolite industry matures, it is reasonable to expect the price to be driven downward.

Conclusions

In summary, nahcolite has the following advantages for use in the proposed baghouse flue gas desulfurization system:

- The material has good operational characteristics in the system.

- There are enormous nahcolite reserves, sufficient for any foreseeable FGD applications.
- The potential selling prices from various sources are reasonable for FGD use.

The major disadvantages of nahcolite with regard to economics are:

- The material is not currently being mined.
- None of the major nahcolite deposits are easily accessible, and large capital investments will be required for mine development.

The preceding disadvantages pose a dilemma for application of nahcolite to FGD:

- The nahcolite lease holders cannot proceed with mine development without an assured market.
- Potential nahcolite consumers cannot commit themselves to the installation of nahcolite-baghouse FGD systems without a guaranteed supply.

Overcoming this situation will likely require either third party--Federal government--involvement or the formation of consortia of either sellers or buyers. In view of the capital requirements and the many technical unknowns involved in opening a mine for nahcolite, a guarantee to the mine developer(s) is not an unreasonable requirement. IRI has stated that a guaranteed market of 500,000 tons/year (450,000 metric ton year) would be sufficient to open a mine, but this figure is dependent on the economic conditions.

WASTE DISPOSAL

Composition and Quantity of Baghouse Wastes

Baghouse filter cake from dry process FGD with nahcolite is composed of spent nahcolite and fly ash. Table 6 compares the composition of nahcolite ore with the expected composition of spent nahcolite. Since the reactions of desulfurization(11)--$NaHCO_3$ conversion to Na_2CO_3, substitution of SO_2 to form Na_2SO_3, and Na_2SO_3 oxidation to Na_2SO_4-- average only 70 percent completion under FGD conditions, both reactants and products are present in the spent

TABLE 6. COMPOSITION OF RAW AND SPENT NAHCOLITE(2)

Raw Nahcolite Constituent	% Dry Weight	Spent Nahcolite Constituent	% Dry Weight
$NaHCO_3$	70	$NaHCO_3$	1
Na_2CO_3	3	Na_2CO_3	17
SUBTOTAL.	73	Na_2SO_3	35
		Na_2SO_4	14
		SUBTOTAL.	67
$MgCO_3$	3	$MgCO_3$-$MgSO_4$	4
H_2O	2	H_2O	2
Other Materials	22	Other Materials	27
$CaCO_3$		$CaCO_3$	
SiO_2		SiO_2	
Al_2O_3		Al_2O_3	
Fe_2O_3		Fe_2O_3	
FeS_2		FeS_2	
Na_2O		Na_2O	
K_2O		K_2O	
Organic Sulfur		Organic Sulfur	
Organic Carbon		Organic Carbon	
TOTAL.	100	TOTAL.	100

nahcolite. Levels of $NaHCO_3$ are low in the spent nahcolite while Na_2CO_3 levels have increased because the formation of Na_2CO_3 is nearly 100 percent and proceeds much faster than subsequent reactions with sulfur.

On a dry weight basis the total percentage of sodium compounds in spent nahcolite is about six percent below the percentage of sodium compounds in nahcolite ore. This weight loss is due to the evolution of CO_2 and H_2O during reaction with SO_2. With a raw nahcolite utilization rate of 30,000 lbs/hr (13,600 kg/hr), 24,700 lbs/hr (11,200 kg/hr) of spent nahcolite are generated.

The main constituents in the fly ash portion of baghouse filter cake are Na_2O, K_2O, CaO, MgO, BaO, carbon, Fe_2O_3, S, Al_2O_3, SiO_2, TiO_2, and P_2O_5. Minor constituents include mercury, zinc, selenium, cadmium, and nickel. With a 4.3×10^5 lbs/hr (2.0×10^5 kg/hr) coal utilization rate, 10 percent ash content in the coal, and 85 percent of total ash released as fly ash, 36,500 lbs (16,600 kg) of fly ash are generated per hour. The total quantity of waste produced per hour from nahcolite FGD is 61,300 lbs (27,800 kg); 40 percent is spent nahcolite, and remaining 60 percent is fly ash.

Table 7 shows the rate of baghouse filter cake generation on an hourly, daily and yearly basis. The uncompacted volume refers to the volume occupied by the baghouse filter cake as it is discharged from the storage silo. The density of the uncompacted waste is 40 lbs/cu ft (640 kg/m^3). The compacted volume is the volume of the waste after compaction in a landfill. Dunlin and Rosar(12) indicate a density of 74.3 lbs/cu ft (1,190.2 kg/m^3) for compacted waste. Each year the planned 500 MW FGD system, then, will generate 214,200 T (190,000 metric ton) of baghouse filter cake with a compacted volume of 5.85 million cu ft (163,000 m^3).

Comparisons of Wastes from Dry Process FGD with Nahcolite and Wet Process FGD with Alkali Scrubbing

The study of disposal of wastes from dry process FGD with nahcolite injection began with a study of available techniques for disposal of wastes generated by wet process FGD systems as well as research and development on disposal of wet process wastes. The purpose of the review of wet process waste disposal was to identify technologies which could be applied to the disposal of the baghouse filter cake, to determine areas where the problems of disposal for the two types of wastes were similar and dissimilar, and finally to use the degreee of environmental protection achieved by wet process disposal systems as an indication of the degree of environmental protection needed for the dry waste.

One important difference in the waste generated by wet and dry process FGD systems is the water content. The dry process waste contains a maximum of two percent water while the water content of the wet process scrubber sludge is approximately 50 percent. As shown in Table 8, the wastes also differ in solubility. Scrubber sludge from lime/limestone systems is primarily calcium sulfate with a relatively low solubility of 2 g/l. Approximately one percent of the sludge is $MgSO_4$, however, which is highly soluble (260 - 700 g/l). The solubility of sodium compounds in spent nahcolite varies from 48 g/l to 700 g/l. Although dry material is more easily disposed of on land, this is balanced by the fact that the sodium compounds of the dry process waste are approximately 100 times more water soluble than the calcium compounds.

The quantities of waste generated in the dry FGD system are greater than quantities produced in lime/limestone scrubbing systems primarily because the products of desulfurization as well as fly ash are contained in the waste. In many lime/limestone systems fly ash is removed by electrostatic precipitation prior to desulfurization. The two wastes are disposed of separately. In dry process FGD with nahcolite, fly ash and spent nahcolite are not separated, and the entire material must be disposed of according to the requirements of the sodium constituents. On a dry weight basis the lime/limestone scrubbing wastes are approximately 40 percent of the dry process baghouse wastes. On a wet weight basis lime/limestone scrubbing wastes are much closer in weight to dry process wastes due to the high water content of the scrubber sludge.

Both lime/limestone scrubber sludge and baghouse wastes contain heavy metals. This is due to heavy metals in the fly ash. The heavy metal content is higher in dry process flue gas desulfurization due to the higher (60 percent) fly ash content.

TABLE 7. RATES OF WASTE GENERATION

Unit of Time	Baghouse Waste Wt. Generated	Uncompacted Volume	Compacted Volume
Hour	30.6 T (27.8 metric ton)	1,530 cu ft (43.3 m^3)	820 cu ft (23 m^3)
24-Hour Day	735 T (667 metric ton)	36,700 cu ft (1,000 m^3)	19,800 cu ft (560 m^3)
7,000 Hr. Year	214,200 T (194,300 metric ton)	10,710,000 cu ft (300,000 m^3)	5,770,000 cu ft (163,00 m^3)

TABLE 8. SOLUBILITIES OF SOME WASTE MATERIALS IN SPENT NAHCOLITE AND SCRUBBER SLUDGE(9)

Waste Material	Compound	Solubility (Grams/Liter of Water)
Spent Nahcolite	$NaHSO_3$	Very Soluble
	$NaSO_3$	125
	$Na_2SO_3 \cdot 7H_2O$	328
	Na_2SO_4 (Thenardite)	48
	$Na_2SO_4 \cdot 7H_2O$	195
	$Na_2SO_4 \cdot 10H_2O$ (Mirabilite)	110
	$MgSO_4$	260
	$MgSO_4 \cdot 7H_2O$ (Epsomite)	710
Lime/Limestone Scrubber Sludge	$CaSO_4 \cdot 2H_2O$ (Gypsum)	2.4
	$MgSO_4$	260
	$MgSO_4 \cdot 7H_2O$ (Epsomite)	710

Although relatively little research has been conducted on disposal of spent nahcolite and fly ash, research on methods for disposal of sludge from lime/limestone FGD has been under way for several years. These procedures were reviewed for their potential application to spent nahcolite/fly ash waste disposal and to determine the degree of environmental protection that would need to be achieved in disposal of spent nahcolite/fly ash. (For a summary of spent scrubber sludge disposal research and development, the reader is referred to the 1974 and 1976 Proceedings: Symposium on Flue Gas Desulfurization(13,14).) From this review it was determined that very little information from scrubber sludge disposal techniques is directly applicable to disposal of dry process wastes. Insolubilization and disposal techniques for scrubber sludge have been developed in response to three properties:

- High water content (50 percent)
- Difficulties in dewatering
- Relatively low solubility of $CaSO_4$

The low water content (<2 percent) and high solubility of spent nahcolite preclude the use of most techniques for scrubber sludge.

Also, the degree of environmental protection achieved by various disposal processes for wet scrubber sludge was not a useful guideline for spent nahcolite/fly ash disposal. At the present time, research, development, and demonstration of several scrubber sludge disposal techniques is being carried out. A review of these techniques and discussions with experts in scrubber sludge disposal(15) indicated that no consensus has been reached on what degree of environmental protection is acceptable.

Specific disposal techniques compared included land disposal, ocean disposal, mine replacement, and insolubilization techniques to reduce the concentration of leachate residuals. The associated land fill designs generally allow for some leachate migration from the site. Because of the high solubility of the waste, complete containment is required for spent nahcolite/fly ash wastes.

Ocean disposal has been considered for both scrubber sludge and baghouse filter cake. Jones(15) identified four environmental concerns related to ocean disposal of scrubber sludge: 1) ocean floor sedimentation--Particle sizes of scrubber sludge are much smaller than the naturally occurring coarse-grained sand particles which are most conducive to the marine life found on the continental shelf; 2) an increase in levels of suspended solids in the water column; 3) sulfite toxicity and COD--COD is created by oxidation of sulfite in the waste to sulfate, and 4) the presence of trace elements such as mercury, zinc, selenium, cadmium, and nickel in concentrations in excess of acceptable levels for the marine environment.

For nahcolite/fly ash wastes ocean floor sedimentation would also be a problem. The sediment would largely reflect the small particle size of the fly ash component; the sodium compounds in spent nahcolite are expected to dissolve quickly.

The possible increase in suspended solids in the water column near the disposal area would also be due primarily to fly ash. With the small particle sizes of fly ash, some difficulties in settling would be expected. The problem of suspended solids would be about equally significant to both scrubber sludge and baghouse filter cake.

The problems of sulfite toxicity may be more significant with spent nahcolite/fly ash waste. Approximately 15 percent of the baghouse filter cake will be sodium sulfite. Trace element contamination in both types of wastes is primarily due to fly ash.

When comparing ocean disposal of wet process and dry process FGD wastes, the environmental problems are similar, although the sulfite toxicity is a more significant problem in the dry process waste. Sulfite content may be reduced by increasing the temperature of the flue gas stream and/or by increasing the reaction time(1,8). Further study is needed to evaluate the sulfite toxicity problem and to determine under what conditions, if any, ocean disposal of spent nahcolite/fly ash would be advisable.

The techniques for mine disposal of wet process sludges are significantly different from dry process techniques and offer little useful information. Some of the problems encountered in a wet process system are leachate from the mine disposal site and stability of the material. The techniques for insolubilization of lime/limestone scrubbing waste are also not transferable for dry process waste.

In summary, research, development and demonstration of techniques for disposal of lime/limestone sludges are still being carried out. No set of criteria for disposal, which could potentially give guidance to baghouse filter cake waste, has been devised. No disposal method is considered totally, environmentally, and economically acceptable for scrubber sludges. Therefore, the experience gained in scrubber sludge disposal can only provide some indication of avenues to explore and possible problems to be aware of.

Guidelines for Disposal/Reuse Investigations

The following assumptions were established to provide a framework from which disposal options were assessed: 1) Surface runoff and leachate to groundwater would need to be avoided in any acceptable disposal technique since leachate from the untreated waste would contain high concentrations of sodium compounds, and 2) Due to the nature of the waste few techniques and possibly no technique would be suitable (environmentally and economically) for all areas of the United States, i.e., techniques for disposal of spent nahcolite/fly ash tend to be site specific.

Waste Disposal Options - General

Open Dumping, Discharge to Surface Streams, and Deep Well Injection--

Open dumping and discharge of wastes to surface streams were considered inappropriate for disposal of nahcolite/fly ash. Due to the high solubility of the waste material, levels of dissolved solids in leachate would exceed existing water quality standards.

Deep well disposal has been used for several types of waste materials. Further research is needed to determine the feasibility of deep well disposal for spent nahcolite. In addition to the common problems of deep well disposal--creating high subsurface pressures, and controlling the migration of injected wastes--problems specific to spent nahcolite deep well disposal include the large volumes of the waste and the need for slurrying.

Disposal in Brine Ponds--

Discharge of spent nahcolite waste materials to brine ponds may be environmentally acceptable. These ponds contain high concentrations of several of the compounds in spent nahcolite. The primary drawbacks to this method of disposal are that it is highly site specific and would probably not be able to accept the typically large quantities of baghouse filter cake generated. Brine pond disposal is probably feasible only for those few small power plants that are located adjacent to a brine pond.

The use of other water catchment basins such as playa lakes and dry desert basins as disposal sites present significant environmental problems. The surface of such basins is generally porous; leachate from spent nahcolite/flyash would potentially increase salinity and trace element levels in the ground water.

Ocean Disposal--

As was discussed in the section comparing scrubber sludge and spent nahcolite disposal, ocean disposal of baghouse filter cake has several environmental problems and requires further study to determine its appropriateness for spent nahcolite/fly ash. The sulfite content is the most significant of the potential problems. Sodium sulfite is toxic to certain marine organisms and also can be expected to reduce oxygen levels in the disposal area through oxidation of sulfite to sulfate. Research would be needed on ways to reduce sulfite concentrations in spent nahcolite. Studies would be needed to determine whether any dispersed disposal techniques would reduce the sulfite toxicity problem.

A second consideration is that the cost of ocean disposal will reflect the large shipping costs associated with the transfer of the spent nahcolite/flyash mixture from its most applicable location to the nearest ocean. At a shipping rate of 1% per ton-mile this would increase the plant operating cost approximately 0.065¢/KW.

Mine Replacement--

Although the high solubility of spent nahcolite/flyash precludes its use as fill material in worked out strip mine areas, replacement of the spent waste in worked out areas of nahcolite ore mines is a promising technique for disposal. The geological characteristics of areas where highly soluble nahcolite ore exists would also provide adequate protection from ground and surface waters. Mine replacement of spent nahcolite may be a means of avoiding subsidence in worked out areas. Economics may be favorable since costs of site preparation will have been borne by the mining operations. Empty cars returning to the mine site for nahcolite ore can be used to transport the spent nahcolite/flyash and thus reduce transportation costs.

Potential problems include the following: 1) With the small particle size of fly ash, dusting will be a significant problem. Techniques such as wetting the waste material may be required to control this problem, but wetting of the waste may not be desirable due to high solubility. 2) The fact that in most cases nahcolite is mined simultaneously with oil shale reduces the feasibility of mine replacement. Nahcolite usually exists as a layer above or below shale, and the cavern created by mining activities may make replacement less feasible. 3) Disposal quantities of spent nahcolite/fly ash are approximately twice the amount of nahcolite ore injected into the FGD system. Additional cars will be required to return the waste to the mine site, and at the most only one-half of the FGD waste could be replaced in the mine. Mine replacement is therefore only a partial disposal solution at best, and some additional means for disposal or reuse of the waste would also be needed.

Land Fill Disposal--

Land disposal is one of the most widely practiced techniques for disposal of waste material. Advantages are generally low costs for disposal where land costs are low. By varying site designs, degrees of protection of surface and subsurface water can be achieved.

For spent nahcolite/fly ash the problems with land disposal are as follows:

1. **Avoiding any contact with water during or after burial in a landfill--Due to the high solubility of the waste, landfill leachate or groundwater in contact with the waste could easily have a dissolved solids content of 50,000 mg/l (see Table 8). Current standards for dissolved solids range from 50-250 mg/l. Also, groundwater contact with the waste or water percolation through the waste would be expected to seriously decrease the stability of the waste.**

To guard against groundwater penetration, the lowest level of the landfill should be approximately 10 feet about the 100 year high level of the water table. During disposal activities contact with rainfall or snow must also be avoided.

Since rainfall runoff may percolate into the waste, the area selected for land disposal of spent nahcolite/fly ash would need to be away from surface streams and free of temporary water tables close to the surface. Surface contours and berms would be required to directed runoff from adjacent areas away from the disposal site.

2. Area runoff--Spent nahcolite/fly ash falling on surrounding areas during waste transportation and disposal can cause area runoff to contain high concentrations of COD and dissolved solids.

3. Impermeable subsurface barriers--It is difficult to ensure the impermeability of subsurface barriers. Concrete, asphalt, butyl rubber, hypalon, vinyl, and clay are among the materials used to line the bottoms of landfills. None of these can seal a landfill indefinitely. For example, asphalt and concrete are sufficiently porous to allow the passage of small quantities of water. A typical rate for concrete is 25 pints/sq yd/yr. Also cracking is a problem with both asphalt and concrete. Rubber, polyethylene and vinyl will effectively prevent leachate from leaving a disposal site; however, the thinness of these materials renders them susceptible to rupture by heavy equipment during the disposal operation.

4. Determining when a subsurface barrier fails--Generally observation wells located around the

perimeter of a site are used to monitor for landfill leachate and thus indicate barrier failure. However, these wells do not monitor the groundwater beneath the disposal site. Much of the area where barrier failure may occur is actually not monitored.

5. Barrier repair--Once a subsurface barrier has failed, repairs are time-consuming and costly because the waste above the repair site must be removed. Due to the large quantities of FGD wastes involved in the present study, barrier failure and repair could be a significant problem.

Although the above problems are important, they are not insurmountable. In certain instances, landfill disposal is an environmentally appropriate technique. Land disposal of spent nahcolite fly ash is best suited to arid and semi-arid areas.

Insolubilization by Aqueous Coprecipitation ("Fersona" Process)--

Dulin et al.(12), have demonstrated the feasibility of insolubilizing sodium sulfate by coprecipitation with acid ferric ions to form the insoluble double salts, natrojarosite and sideronatrite. The double salts produced are less soluble than calcium sulfate and do not tend to form sludges since the precipitate is generally an easily dewaterable granular crystalline solid. The greatest drawback to this technique is the availability of the acidic iron solution for use in the coprecipitation process. Copper smelter/cement copper operations do have an iron-rich barren liquor which can be coprecipitated with spent nahcolite to produce natrojarosite and sideronatrite. Additional sources of acidity and ferric ions may be sulfuric acid from flue gas side streams and iron values in fly ash or bottom ash, waste acid mine water such as black water or gob leachate which typically contains both sulfuric acid and ferric ion values, and other waste from various industrial processes.

Insolubilization by Dry Sintering ("Sinterna" process)--

Dulen et al.(12), report that the solubility of spent nahcolite/fly ash waste can be reduced to below that of gypsum by pelletization of the waste material and sintering. The typical solubility range is 50-1000 mg/l of sodium. The solubility of the sintered pellet increases as the proportion of spent nahcolite increases. Dulin et al.(12), report that no trace elements were detected in the leachate. Some evolution of SO_2 does occur during sintering; quantities released have ranged upward from 0.4 percent of the starting materials. The sintered product has approximately twice the density of compacted baghouse filter cake, i.e., 155 lbs/cu ft (2483 kg/m^3).

Disposal of the relatively insoluble sintered pellets would probably require less stringent measures for leachate control although some protection from the environment would be needed. The material would probably require contouring to facilitate runoff and a final surface cover of clay. Vegetation could then be planted on the surface and the land could be recovered for a multitude of uses such as parking lots, parks and buildings. The landfill area should be protected from groundwater. The sintered pellets would also be suitable for disposal in strip mining sites. Possibilities for use as a product include concrete aggregate and road bed base.

Except for economic constraints, land disposal of sintered pellets appears to be widely applicable. In arid areas the sinterna process may be non-competitive with disposal of untreated baghouse waste in clay isolation cells. Estimates in 1976 of costs for the Sinterna process on an annual basis are $8.50/T. This figure does not include costs for waste hauling and disposal, and for the minimal amount of land contouring that would probably be required.

Methods for Recovery/Use as a Product--

In each of these cases below, the applicability of the spent nahcolite/fly ash is limited but appears promising in certain site-specific situations.

Pulp and Paper Mills--In some cases spent nahcolite may be used by a craft paper plant for bleaching purposes(12). Because the filter cake is fly ash contaminated and because the market is small only a small part of the dry FGD waste could be used in this way.

Glass Manufacturing--The spent nahcolite and glass batch fines collected in a baghouse of a glass manufacturing plant may be recycled to the glass batch. Nahcolite gangue material (silica, alumina, Ca/$MgCO_3$, potassium, feldspars, and analcime) is compatible with glass batch formulations. Further, glass manufacturing requires sodium sulfate. As glass manufacturing facilities change to coal-fired furnaces this process will become more attractive. The flue gases would contain more SO_2 and ash fines if coal were to be used. The process is potentially a closed loop system since the wastes from SO_2 emissions are directly used in glass manufacture. Glass manufacturing facilities can accept more spent nahcolite/fly ash than would be generated on site; however, a consistent composition of the material would be required for glass batch formulation.

Integrated Pyrometallurgical Copper Smelter/Cement Copper Facility--In an integrated pyrometallurgical copper smelter/cement copper operation spent nahcolite would be used to reduce the iron content of one of the process streams and in the process would be converted to less soluble material. Specifically, spent nahcolite is forwarded to the cement copper operation where it is reacted under oxidizing conditions with the ferrous ions in the "barren liquor." The sodium compounds in spent nahcolite are coprecipitated as the insoluble double salts sideronatrite and natrojarosite. The liquor, now free of ferrous ion, is returned to the leach dumps. Using sodium sulfite/sulfate waste in this manner would actually create a leaching liquor of greater reaction efficiency due to the lower concentration of iron salt in the solution. From the standpoint of the baghouse waste material the conversion products, sideronatrite and natrojarosite, are disposed of via landfill.

Waste Disposal - Specific

From the general review of waste disposal methods, it was determined that most methods are site specific. This led to a study of the disposal methods that could be used in the mid-western region of the U.S. where construction of the test facility is recommended. Climate, geology, and facility size are important determinants of appropriate disposal methods.

The Mid-west is typically a semi-arid region with rainfall of 20 inches (0.5 m) or less per year and a water table at 50 ft (15 m) or more below the surface. Common plant communities include short grass prairie and creosote bush. Soil types range from sandy to clay; the soil varies from slightly to highly alkaline.

Of the disposal methods discussed in the previous section, the following were considered most appropriate for the type of waste and the region:

- "Sinterna" process and land disposal
- Land disposal without insolubilization
- Mine replacement
- Integrated copper processing

Of these, burial in clay isolation cells and insolubilization via briquetting and sintering ("Sinterna" process) followed by land burial were given in-depth consideration. Although promising, use in an integrated copper processing facility was too site specific to be given further study. It was felt that not enough was known about nahcolite mining methods, distance of test facility from mine mouth, and so forth to allow further consideration of mine replacement at the present time.

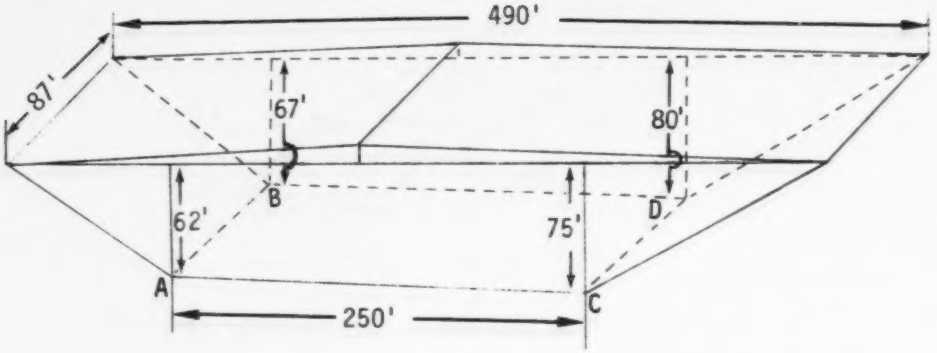
Burial in Clay Isolation Cells--

The general discussion of disposal methods lists several potentially serious environmental problems which must be addressed if land burial of waste material is planned. Svec, et al.(16), EPA(17), and TRW(18), along with many others have documented damage to land, surface water, and ground water due to the improper use of land disposal methods. Since no commonly used landfill design adequately addressed all potential environmental problems, a new design is suggested for disposal of spent nahcolite/fly ash.

Site Requirements--Because of the high solubility of the spent nahcolite/fly ash waste material it is imperative that the water table be at least 10 feet (3 m) below the bottom surface of the landfill; this level for the water table should represent the one hundred year high point. The site must also be free of surface streams flowing on either a continuous or an intermittent basis. Although soil at the disposal site can in fact be of any kind, fill will be located in a clay soil.

Recommended Landfill Design--The only suitable land disposal method is one that avoids ground water and surface contamination by keeping the waste from contacting water during and after disposal operations. If this difficult task is accomplished, there will be no leachate generated and hence the waste will be successfully isolated. The recommended landfill design does not employ a subsurface barrier due to difficulties in maintenance and due to difficulties in determining when a barrier has been penetrated. Instead of a subsurface barrier, a barrier across the surface of a finished cell has been utilized. A surface barrier allows for routine and frequent inspections for barrier failure, and any needed repairs can be made quickly and economically without disturbing the waste. Replacement of the barrier can take place as needed and the perpetual maintenance necessary to insure that the waste does not contaminate ground water or surface water is possible. The landfill can also be located in virtually any kind of terrain, i.e., not necessarily in clay since no leachate will be generated. The landfill will also maintain its integrity even during mild geological disturbances as long as a major shift in the water table does not occur. This technique for land disposal successfully addresses each of the central environmental problems for land disposal stated in the previous sections, provided that this landfill design is accompanied by land siting which provides protection from ground water and temporary perched water tables.

To keep the waste dry during disposal operations, the clay isolation concept was utilized. We have used a slight modification of the trench method(19) although the basic concepts of the design could also be applied to the area method or to an area-trench combination method. The trench method was chosen because the specific surface contours of the disposal site are not known. As is shown in Figure 15 the floor of the excavation site is tilted to one corner. This is to direct rainfall to one point where the water can be pumped to an evaporation pond. (Periodically, the residue will be removed from the pond and replaced in the disposal site.) The difference between points A and D is 18 ft (5.5 m). During disposal operations this slope will be maintained. When a cell is completely filled, final grading will be done to achieve the contour shown.



Scale: 1/10" = 10'

FIGURE 15. DIMENSIONS OF FINISHED LANDFILL CELL

The basic building block for the landfill is the clay isolation unit. The waste material generated from one day's operation will be transported to the site, compacted, and covered with a one foot layer of clay which is also compacted. This forms the clay isolation unit. Thus, each day the soluble sodium sulfite is encapsulated in an essentially impermeable cell of clay. It is important that the clay soil material have low permeability so that any rainfall will run off of the sloping surface rather than penetrate the material. As subsequent units are added to form a layer, the height of a unit constitutes a layer height.

When the sodium waste is delivered to the landfill operation by truck, it is dumped at the toe of a previously finished cell. A crawler dozer spreads the spent product up at a 30 degree slope to the previous cell. The thickness of the spread layer is kept less than 1-1/2 ft (0.3-0.5 m) in order to obtain better compaction by the tractor dozer. At least five passes by the tractor are required over the spread material in order to assure greater volume reduction. For each truckload, the material is handled in the same manner. By the end of the work day, the unit (containing 735 T (667 metric ton) of waste) will have a trapezoid cross section approximately 13 ft (4.0 m) high, 18 ft (5.5 m) wide, and 87 ft (26.5 m) long, as shown in Figure 16. The cell is covered with 1 ft (0.3 m) of compacted clay-like material.

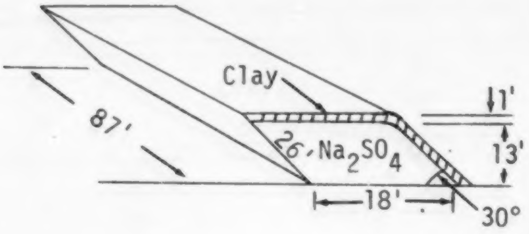


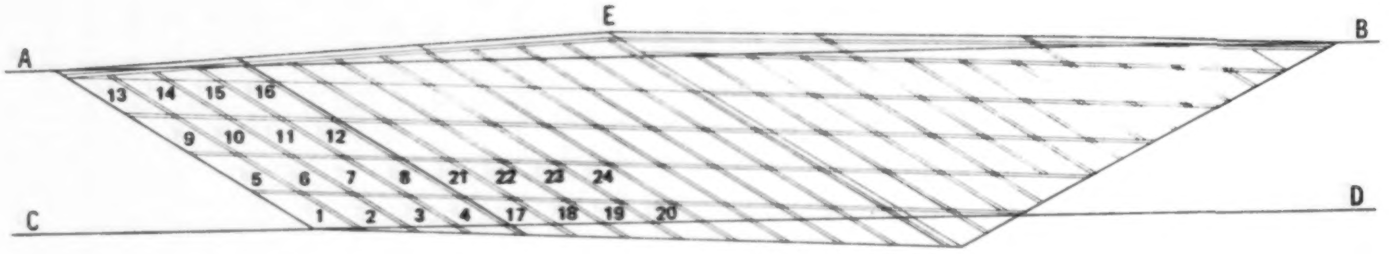
FIGURE 16. TYPICAL CLAY ISOLATION UNIT(12)

For one landfill cell, a total of four layers are employed, and the final cover for the top will be a compacted thickness of 3 ft (0.9 m) of soil with the butyl liner installed above the clay. A butyl liner was selected because of its guaranteed resistance to the effects of weathering over a minimum of 20 years.

Because of the importance of keeping the material dry, disposal operations will cease during rainfall. A silo with 8 days storage capacity has been planned so that ample time will be allowed for covered storage during rainfall and the subsequent drying periods.

Figure 17 shows a cross-section along the length of one finished landfill cell. Each of the trapezoidal shaped squares represents one day's operation, one unit. Line $\overline{AB}$ represents the flat surface of the surrounding area and shows that the surface of the landfill is sloped to a center point, E. Line $\overline{CD}$ is parallel to the surface and indicates the slope of the floor of the landfill. For every 20 ft (6.1 m) of length, there is a 1 ft (0.3 m) downward slope of the landfill floor. Also, the floor of the landfill is tilted 1 ft (0.3 m) for every 20 ft (6.1 m) of width.

The sloping floor somewhat increases the complexity of this landfill; however, this is simplified by the fact that once the slope is achieved, the contour is maintained throughout the first three levels of filling. The



Scale: 1/16" = 4'

FIGURE 17. LENGTHWISE VIEW OF FINISHED LANDFILL CELL

tilting of the floor is then changed in the fourth and final level to create the final surface contour as shown in Figure 17.

There are several ways that each landfill cell could be filled. For instance, the first level could be completely filled before the second layer is added. This method of filling is not recommended because, to the extent possible, the rainfall should collect in an area where there is no fill material. The 4-day sequence plan shown in Figure 17 has been chosen. This is based on placing 4 cells across the bottom of the landfill and then filling above these 4 areas for 4 subsequent, 4-day periods. At the end of slightly over 2 weeks then the first subcell of the landfill will be completely filled with waste material and final contour developed. The next 4 days of operation will then be placed on the bottom of the landfill and the process repeated. This will continue until the entire site has been filled.

When a site has been filled and the final contouring accomplished, the lining material of butyl rubber will be installed across the surface. The lining material will be secured by sandbags placed at 3 ft (0.9 m) intervals across the surface. The lining material will be inspected as needed for signs of failure. The expected lifetime for a butyl liner is a minimum of 20 years. Sandbags and liners will be replaced as needed.

One landfill cell will require approximately 1 acre (0.4 hectare) of land. About 71,000 tons (64,400 tonne) of spent nahcolite/fly ash can be placed in one landfill cell. This represents approximately 1/3 of the annual amount of waste generated by the 500 MW test facility. Therefore, 3 acres (1.2 hectares) of land per year would be required for land disposal of the spent nahcolite/fly ash. A minimum of 60 acres (24 hectares) would be required for a plant with an expected 20 year life.

Surface Contours at the Disposal Site--To prevent the entry of runoff from adjacent areas into disposal sites, a protective berm is constructed around the perimeter. The area outside of the berm is graded as needed to direct runoff away from the disposal area. Within the berm, the surface would be contoured to direct runoff water to the evaporation pond. The evaporation pond would be of sufficient size to accept all water falling to the area.

Environmental Assessment--The surface sealing method for disposal of spent nahcolite/fly ash eliminates the problems of ground water and surface water contamination. Two environmental effects remain. First, the water that falls onto the disposal area will be prevented from reaching the water table and to that extent will reduce the level of the water table. Because the area involved is relatively small, this effect is expected to be insignificant. Second, the land will be unavailable for use for other purposes. The waste would be easily retrieved for conversion to other useful products, however.

Insolubilization by Pelletization and Sintering - "Sinterna" Process--

As described in the general discussion of disposal methods, the solubility of spent nahcolite/fly ash waste material may be significantly decreased by the application of the "Sinterna" process. The solubility is decreased approximately three orders of magnitude; the solubility range of the sintered pellets is 50 - 1,000 mg/l. In the arid/semi-arid area under consideration in this study where the material could be disposed of on the surface in large piles with runoff directed to a large evaporation pond. Alternatively, the sintered pellets could be disposed of by more conventional landfill techniques. Daily soil cover would still be necessary to minimize water penetration and dusting.

Economics

In determining the cost for disposal of spent nahcolite/fly ash waste material, it was assumed that the power plant would own the land needed for disposal and that the roads leading to the disposal area are in place.

Land Disposal in Clay Isolation Cells-

In order to determine costs for land disposal, the following landfill operating conditions were assumed:

- A disposal site location within 15 miles (24 kilometers)
- A 7 day, 8 hour per day operating schedule
- 6 hours of average daily operation for each vehicle

Table 9 shows the breakdown of capital and operating costs. The capital costs are low since land and road costs are not included and since

TABLE 9. CAPITAL AND OPERATING COSTS FOR LAND DISPOSAL*

Item	Capital Costs (Actual Dollars)	Operating Costs (Actual Dollars)
Site Preparation		
Fence	24,200	
Administration Building	25,000	
Maintenance Building	25,000	
Well	5,000	
Pump	2,000	
Fuel Tank and Pump	5,000	
SUBTOTAL	86,200	
Waste Hauling		
Tractor Rentals, 5 Ea.		108,000
Trailer Rental, 4 Ea.		27,700
Maintenance		10,000
Fuel		23,100
Personnel (8), including Fringes @ 30%		178,000
SUBTOTAL		346,800
Excavation and Filling		
Equipment Rental		388,800
Fuel		35,000
Maintenance		12,000
Personnel (18), including Fringes @ 30%		400,000
SUBTOTAL		835,800
Liner		
Installation	1,200	7,000
Butyl Rubber Liner Material		79,700
Contingency, 20 Year Basis		30,000
SUBTOTAL	1,200	116,700
TOTAL CAPITAL COSTS	87,400	
TOTAL ANNUAL OPERATING COSTS		1,299,300

*Based on a 500 Mw plant, 7000 hrs/year operation, burning 1,0% sulfur coal with a heating value of 10,500 Btu/lb.

all trucking and excavating equipment costs are based on lease arrangements. Of the total cost per ton for disposal, only $0.02 is due to capital costs.

The operating costs are due primarily to equipment, manpower, and liner costs. The first step in determining equipment costs was to determine the size of equipment needed. One of the determining factors for equipment size was the storage silo. It was estimated that a minimum of 10 minutes would be needed for bringing a truck in, loading, and pulling away from the storage silo. A 30 percent factor was then added to total 13 minutes. Assuming a 6 hour day for hauling, 28 trailers could be loaded each day. Based on an average daily generation of 750 T (680 metric ton), each load would weigh 27 T (24 metric ton). Since the 15 mile (24 kilometers) round trip to the landfill was estimated to take 50 minutes, 4 waste hauling vehicles were used. Equipment costs were determined by receiving quotes from venders. Large scale, off the road equipment is used for waste hauling. As is shown in Table 10, 4 tractor-trailer waste haulers (plus 1 backup tractor), 1 road grater, 4 crawler tractors, and 1 self-loading scraper are required.

Operation of this equipment on a 7-day basis, direction of vehicles, maintenance, and supervision requires 26 people. Labor rates have been based on cost escalations from the TVA study of FGD processes(2). The methods by which the escalation was derived are explained in the Economics Section.

Liner costs include price quotes for the actual liner and a contingency set aside for maintenance and liner replacement after the power plant ceases operation.

The annual operating costs for disposal of baghouse filter cake from the 500 MW test facility are $1,300,000. The cost per ton in 1977 dollars is $6.10.

"Sinterna" Process of Insolubilization--

In-depth costs were not developed in this study for the "Sinterna" process. Quotes received from Industrial Resources Inc. estimate a cost per ton of $8.40 (June 1976 dollars) for briquetting and sintering. This does not include the additional cost for disposal which we estimate to be $5.60/ton if land burial is used and $3/ton if surface disposal and evaporation ponds are used.

ECONOMICS

The choice of FGD systems for use at any particular site is principally dependent on the relative costs for the various available systems. These costs will vary for specific installations in accordance with unit size, combustion coal analysis, and emissions regulations. Both capital costs and annual operating costs are considered in this evaluation. Representative costs for other flue gas desulfurization schemes are included for comparison purposes.

Methodology

This section develops the basic tools and techniques used in evaluating the capital and operating costs of the dry sorbent baghouse system. Equipment has been broken down into categories and each category has been assigned a specific set of indices relating it to various inflation indices. A method of combining the various economic factors, such as equipment costs, installation labor, contractor fee, is presented and applied to the dry sorbent baghouse system.

Cost Indices--

Cost indices are a means of estimating the cost of a certain piece of equipment at any desired point in time by knowing the equipment cost at some other point in time. Since they are an invaluable aid in predetermining planned plant costs, Chemical Engineering publishes monthly tabulations of plant cost indices which are subdivided into several equipment categories.

Price indices are given for selected group headings such as fabricated equipment, pumps and compressors, construction labor, and are a numerical representation of the relative increase of each division for which an index is given referenced to a preset base period (which, in this case, is 1957-1959).

To determine an estimated cost for any item of equipment, a previous cost for a similar item must be known. The previous cost is divided by the cost index for the time period during which that cost was incurred and the result is multiplied by the cost index for the time period for which the estimated cost is desired.

Item Categorization--Each item of equipment associated with the dry sorbent FGD system has been classified into specific categories associated with the economic indices. This has allowed the effective comparison of individual costs with those of other FGD processes. Categories used for this assessment are:

Equipment

Materials Handling

Excavation

Foundations

Engineering

Contractor

Contingincy

Other Construction Costs

Estimating Factors--

Process costs have generally been derived from an extrapolation of material acquisition costs, based on a known or assumed relationship between the various individual costs associated with a given equipment system. This relationship will differ depending on the specific type of equipment considered, the amount of new or uncommon technology, and the labor intensiveness of the installation.

Cost Estimation Technique--Individual equipment costs are combined with associated design, procurement, and construction costs to develop a total module cost. The total module costs are then added to provide a total system cost. When not directly available, each of the associated design, procurement, and construction costs are estimated from the equipment costs in accordance with the method of K.M. Guthrie(20). Figure 18 depicts the methodology used for the establishment of these costs. Factors k_1 through K_7 are coefficients relating the various components of a module cost and are individually established for each equipment classification. A brief description of each component cost is as follows:

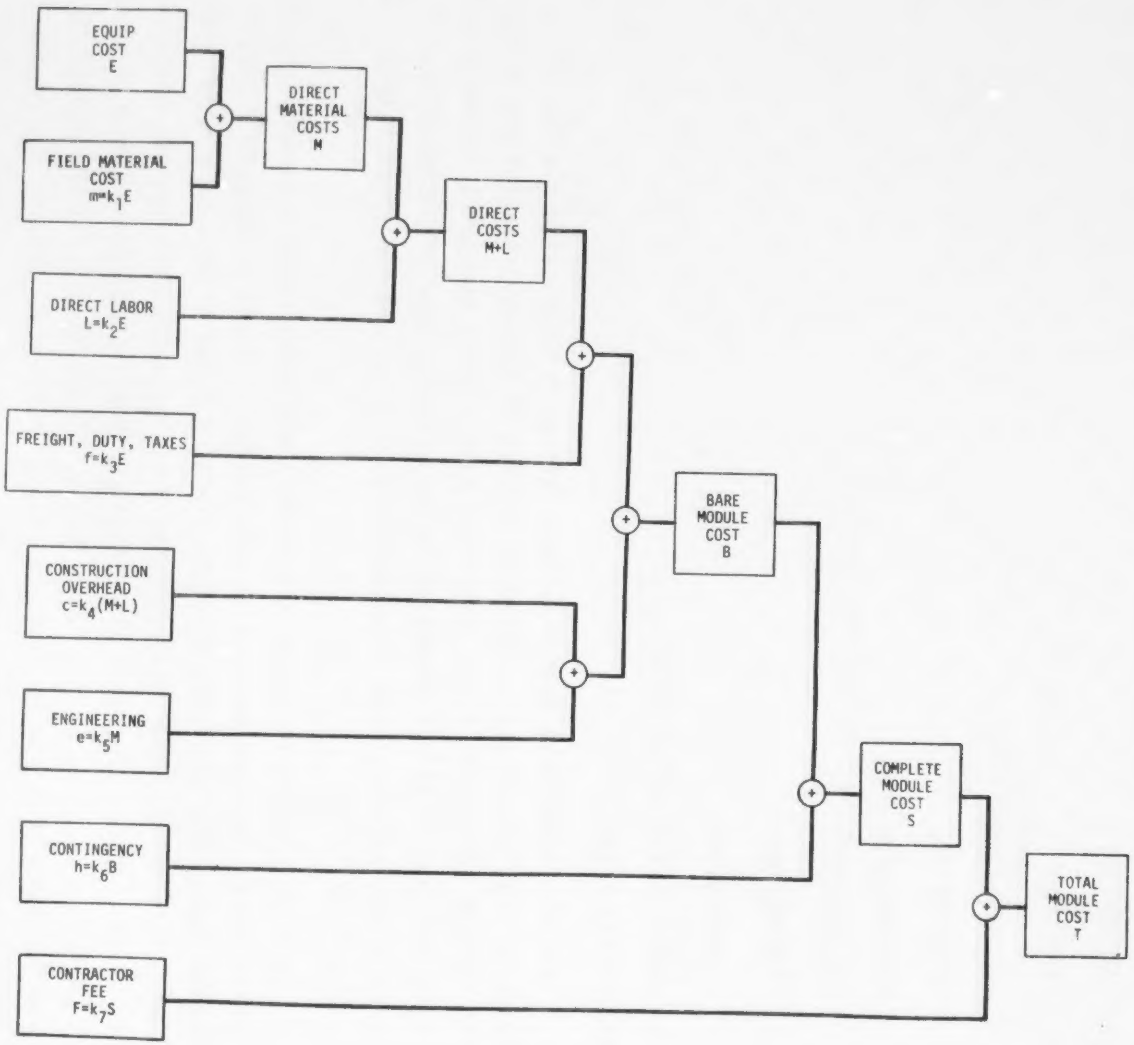


FIGURE 18. COST ESTIMATION METHODOLOGY

Equipment Cost (E) - The purchase price of the major item(s) of equipment that contribute to an individual cost module.

Field Material Costs (m) - The purchase price of the total amount of materials necessary for the erection and operation of the individual cost module, except those contained in (E).

Direct Labor (L) - All field labor, excluding fringe benefits, assuming an average mix of crafts, associated with the individual cost module.

Freight, Duty, Taxes (f) - Freight charges, taxes, and licenses are based on normal rates for shipment within the continental U.S.

Construction Overhead (c) - This component includes labor fringe benefits, FICA, Federal and state unemployment insurance, workmen's compensation, supervisory personnel, temporary construction facilities, heavy equipment leasing, purchase small tools, and miscellaneous support equipment and services.

Engineering (e) - Engineering services directly attributable to the individual cost module plus a proportionate amount of the total project engineering not directly attributable to any individual cost module. Engineering also includes costs associated with procurement, drafting, and all indirect and overhead office costs associated with the project engineering.

Contingency (h) - The cost of unlisted items and insufficient scope definition expected to be incurred in the individual cost module.

Contractor Fee (F) - The portion of the total contractor fee attributable to the individual cost module.

Index Assignment--To allow an updated estimation of module costs, we have developed various "inflators" which are an indication of the increase in the individual component costs which make up the module cost. These "inflators" are computed for such classifications as engineering rates; freight, duty, and taxes; field-erected steel structures; and rentals, and are developed using Chemical Engineering price indices, the Consumer price index and increases in FICA rates as determined by the U.S. Department of Health, Education, and Welfare. Table 10 gives the breakdown of the "inflators" calculated as well as the means of determining their numerical value.

Estimation Coefficients--After having generated the "inflators" to update the component costs to 1977 dollars, it is possible to establish estimation coefficients as discussed in the Cost-Estimation Technique section which reflect conditions in February 1977. Since the Chemical Engineering plant cost indices are used to update several historical cost estimations to present-day dollars (i.e., February 1977 dollars), it is necessary to ascertain what effect each of the given indices will have on the aforementioned headings.

In order to develop a cost index for each category, an estimate of the percent effect each plant cost index has is shown in Table 11. These coefficients are established by discounting or appreciating the applicable coefficients (based on the piece of capital equipment being analyzed) in reponse to the change in the value of the inflators being used. Sample calculations for the determination of the estimation coefficients for the system ID fans are given in Appendix A.

Using the same general procedure as outlined by the above sample calculations, estimation coefficients are derived for the various modules established for the nahcolite/baghouse process and are given in Table 12.

Capital Costs

The costs associated with the design, procurement, erection and start-up of a dry sorbent baghouse flue gas desulfurization system employing the previously described process parameters has been established as a base case. Individual equipment costs are identified, module costs are calculated in accordance with the above procedure, and from these the total project cost is determined. The effect on this total cost due to the differences in design anticipated for specific installations are then evaluated.

TABLE 10. CALCULATION OF INFLATORS

1. LABOR RATES

Using construction labor indices of *Chemical Engineering*:

July, 1968 = 119.8

February, 1977 = 176.1

$$\longrightarrow \text{inflator} = \frac{176.1}{119.8} = 1.4699$$

$$\boxed{\text{INFLATOR: } 1.4699}$$

2. ENGINEERING RATES

Using engineering and supervision indices of *Chemical Engineering*:

July, 1968 = 109.1

February, 1977 = 154.4

$$\longrightarrow \text{inflator} = 1.4152$$

$$\boxed{\text{INFLATOR: } 1.4152}$$

3. FREIGHT, DUTY, TAXES

Using consumer price indices as determined by U.S. Bureau of Labor Statistics:

July, 1968 = 100.0 (made base year)

February, 1977 = 170.3

$$\longrightarrow \text{inflator} = \frac{170.3}{100.0} = 1.7030$$

$$\boxed{\text{INFLATOR: } 1.7030}$$

TABLE 10. (Continued)

4. FIELD-ERECTED STEEL STRUCTURES

Using the following price indices from *Chemical Engineering* in proportion given:

Pipe, valves, fittings	20%
Fabricated equipment	40%
Structural supports	40%

1968 index:

$$(.2)(116.5) + (.4)(109.2) + (.4)(105.2) = 109.06$$

1977 index:

$$(.2)(238.1) + (.4)(208.5) + (.4)(219.2) = 218.70$$

$$\longrightarrow \text{inflator} = \frac{218.70}{109.06} = 2.0053$$

INFLATOR: 2.0053

5. SOLIDS-HANDLING EQUIPMENT

Using the following price indices from *Chemical Engineering* in proportion given:

Pipe, valves, fittings	10%
Process machinery	30%
Pumps, compressors	30%
Structural supports	30%

1968 index:

$$(.1)(116.5) + (.3)(111.7) + (.3)(115.3) + (.3)(105.2) = 111.31$$

1977 index:

$$(.1)(238.1) + (.3)(206.1) + (.3)(234.3) + (.3)(219.2) = 221.69$$

$$\longrightarrow \text{inflator} = \frac{221.69}{111.31} = 1.9916$$

INFLATOR: 1.9916

(Continued)

TABLE 10. (Continued)

6. PUMPS, MOTORS

Using pumps and compressor price indices from Chemical Engineering:

July, 1968 = 115.3
February, 1977 = 234.3

$$\longrightarrow \text{inflator} = \frac{234.3}{115.3} = 2.0321$$

INFLATOR: 2.0321

7. CONSTRUCTION OVERHEAD

Using increase in consumer price index (CPI) as determined by U.S. Bureau of Labor Statistics, increase in FICA (SS) as determined by U.S. Department of Health, Education, and Welfare, and increase in construction cost price index (CCPI) from Chemical Engineering:

(SS)	FICA	7%
(SS)	Unemployment insurance	7%
(CPI)	Employment benefits	16%
(CPI)	Insurance	20%
(CCPI)	Construction equipment	50%

$$(.07)(0.5395) + (.07)(.5395) + (.16)(.703) + (.20)(.703)$$
$$(.50)(1.1069) = 0.8821$$

$$\longrightarrow \text{inflator} = 0.8821$$

INFLATOR: 0.8821

(Continued)

TABLE 10. (Continued)

8. RENTAL

Using construction cost price indices from Chemical Engineering:

July, 1968 = 1165.5

February, 1977 = 2508.9

$$\longrightarrow \text{inflator} = \frac{2508.9}{1165.5} = 2.1526$$

$\boxed{\text{INFLATOR: } 2.1526}$

9. GENERAL EQUIPMENT COSTS

Using equipment, machinery, supports, price indices from Chemical Engineering:

July, 1968 = 110.9

February, 1977 = 213.7

$$\longrightarrow \text{inflator} = \frac{213.7}{110.9} = 1.9270$$

$\boxed{\text{INFLATOR: } 1.9270}$

10. CONTINGENCY

Using following price indices from Chemical Engineering in proportion given:

Construction labor	10%
Buildings	10%
Engineering and supervision	10%
Fabricated equipment	10%
Process machinery	10%
Pipe, valves, and fittings	10%
Process instruments	10%
Pumps and compressors	10%
Electrical equipment	10%
Structural supports and miscellaneous	10%

(Continued)

TABLE 10. (Concluded)

10. CONTINGENCY (Continued)

1968 index:

(.1)(119.8) + (.1)(114.9) + (.1)(109.1) + (.1)(109.2) + (.1)(111.7) + (.1)(116.5) + (.1)(120.1) + (.1)(115.3) + (.1)(91.1) + (.1)(105.2) = 111.29

1977 index:

(.1)(176.1) + (.1)(194.4) + (.1)(154.4) + (.1)(208.5) + (.1)(206.1) + (.1)(238.1) + (.1)(199.6) + (.1)(234.3) + (.1)(154.9) + (.1)(219.2) = 198.56

$$\longrightarrow \text{inflator} = \frac{198.56}{111.29} = 1.7842$$

INFLATOR: 1.7842

11. CONTRACTOR

Using following price indices from Chemical Engineering in proportion given:

Construction labor	50%
Engineering and supervision	50%

1968 index:

(.5)(119.8) + (.5)(109.1) = 114.45

1977 index:

(.5)(176.1) + (.5)(154.4) = 165.25

$$\longrightarrow \text{inflator} = \frac{165.25}{114.45} = 1.4439$$

INFLATOR: 1.4439

TABLE 11. ALLOCATION OF CAPITAL EXPENSE ITEMS TO CHEMICAL ENGINEERING PRICE INDICES

CHEMICAL ENGINEERING PRICE INDEX / INVESTMENT INPUT	CONSTRUCTION LABOR	BUILDINGS	FABRICATED EQUIPMENT	PROCESS MACHINERY	PIPE, VALVES, FITTINGS	PROCESS INSTRUMENTS	PUMPS COMPRESSORS	ELECTRICAL EQUIPMENT	STRUCTURAL SUPPORTS AND MISCELLANEOUS	ENGINEERING AND SUPERVISION	TOTAL
EQUIPMENT			60%	20%			20%				100%
MATERIALS HANDLING					20%		50%		30%		100%
EXCAVATION	100%										100%
FOUNDATIONS			50%						50%		100%
ENGINEERING										100%	100%
CONTRACTOR	50%									50%	100%
CONTINGENCY	10%	10%	10%	10%	10%	10%	10%	10%	10%	10%	100%
OTHER CONSTRUCTION COSTS	10%	10%	10%	10%	10%	10%	10%	10%	10%	10%	100%

TABLE 12. ESTIMATION COEFFICIENTS FOR USE IN BASIC CAPITAL COST MATRIX

	k_1	k_2	k_3	k_4	k_5	k_6	k_7
MILL SURGE TANK	0.086	0.007	0.068	0.168	0.096	0.140	0.021
BAGHOUSE PRECOAT INJECTION SYSTEM	0.150	0.360	0.068	0.183	0.097	0.148	0.022
NAHCOLITE INJECTION SYSTEM TO DUCT	0.150	0.360	0.068	0.183	0.097	0.148	0.022
CONVEYING SYSTEM	0.150	0.360	0.068	0.183	0.097	0.148	0.022
CONVEYOR AND FEEDING NETWORK OF MILL	0.150	0.360	0.068	0.183	0.097	0.148	0.022
DUCT SYSTEM*	N.A.	N.A.	N.A.	0.154	0.097	0.148	0.022
SITE STORAGE SILO	0.086	0.007	0.068	0.168	0.096	0.140	0.021
ASH DISPOSAL SILO	0.086	0.007	0.068	0.168	0.096	0.140	0.021
STORAGE DAY TANK	0.086	0.185	0.068	0.176	0.096	0.140	0.021
MILL	0.429	0.106	0.068	0.174	0.086	0.143	0.022
BAGHOUSE	N.A.	N.A.	N.A.	N.A.	N.A.	0.150	0.030
ID FANS	0.256	0.246	0.067	0.177	0.096	0.145	0.021
ASH DISPOSAL SCREW CONVEYOR	N.A.	N.A.	0.068	0.183	0.097	0.148	0.022

N.A. = NOT APPLICABLE

*VALUES DETERMINED BASED ON CONVEYING SYSTEM COEFFICIENTS AND ID FAN COEFFICIENTS

Equipment Costs--

Evaluation of module costs required that the cost of each component that is included in the nahcolite/baghouse system be known or reasonably estimated. To ascertain each individual equipment cost, one of two methods was employed: obtaining costing information from various vendors or utilizing estimates from historical data.

New Equipment Cost Quotes--For the majority of equipment costs used in this study, vendors were solicited to establish costing data for individual pieces of capital equipment. In order to accurately gauge the cost of each piece of equipment, a minimum of two different manufacturers were asked to submit quotes and the more reasonable of the two obtained bids was used.

Vendors, for the most part, represented major manufacturing concerns who have much expertise in the electrical generation, pollution abatement field. They were asked not only to submit bids for the cost of the desired piece of equipment, but to also estimate the cost of erection, cost of ancillary equipment, if any, and any other related costs. In some instances, the quoted price included the cost of engineering, erection, foundations, and the like while, in other cases, only the bare equipment costs were ascertainable.

Estimates from Historic Data--Costs of the major storage tanks (those that are utilized to store the raw, unmilled nahcolite and the baghouse wastes) were developed from estimations based on the TVA study. However, since these costs are given in 1975 dollars, it was necessary to update them to February 1977. This was done by utilizing the applicable plant cost price indices given in Chemical Engineering.

Base Case Evaluation--

From the above equipment costs, an estimate of the individual module costs, which make up the total project costs, can be calculated by use of the capital cost estimating structure as shown in Figure 18. From this matrix, it is possible to determine such capital items as labor costs; freight, duty, and taxes; engineering costs; contingency costs; and other costs that are included in the individual module costs. Once the module costs have been computed, the total capital cost can be estimated. Table 13

TABLE 13. ALLOCATION OF MODULE COSTS, BASE CASE

	NAHCOLITE UNLOADING SECTION	SITE STORAGE SILO	CONVEYORS	SURGE TANK	CONVEYORS AND FEEDING NETWORK OF MILL	MILL	STORAGE DAY TANK
EQUIPMENT COSTS	3,500,000	1,200,000	74,000	50,000	60,000	180,000	38,000
FIELD MATERIAL COSTS	INCLUDED	103,200	11,100	4,300	9,000	77,200	3,300
DIRECT MATERIAL COSTS	3,500,000	1,303,200	85,100	54,300	69,000	257,200	41,300
DIRECT LABOR	INCLUDED	8,400	26,600	400	21,600	19,100	7,000
DIRECT COSTS	3,500,000	1,311,600	111,700	54,700	90,600	276,300	48,300
FREIGHT, DUTY, TAXES	157,800	81,600	5,000	3,400	4,100	12,200	2,600
CONSTRUCTION OVERHEAD	643,300	220,300	20,400	9,200	16,600	48,100	8,500
ENGINEERING	258,800	125,100	8,300	5,200	6,700	22,100	4,000
BARE MODULE COSTS	4,559,900	1,738,600	145,400	72,500	118,000	358,700	63,400
CONTINGENCY	674,900	243,400	21,500	10,200	17,500	51,300	8,900
COMPLETE MODULE COSTS	5,234,800	1,982,000	166,900	82,700	135,500	410,000	72,300
CONTRACTOR	115,200	41,600	3,700	1,700	3,000	9,000	1,500
TOTAL MODULE COSTS	5,350,000	2,023,600	170,600	84,400	138,500	419,000	73,800

TABLE 13. (Continued)

	INJECTION TO DUCTS	BAGHOUSE INJECTION	BAGHOUSE	ID FANS	DUCTS	ASH DISPOSAL CONVEYOR	ASH DISPOSAL SILO	WASTE DISPOSAL
EQUIPMENT COSTS	16,000	16,000	8,500,000	495,000	200,000	25,000	450,000	87,400
FIELD MATERIAL COSTS	2,400	2,400	INCLUDED	126,700	INCLUDED	INCLUDED	38,700	INCLUDED
DIRECT MATERIAL COSTS	18,400	18,400	8,500,000	621,700	200,000	25,000	488,700	87,400
DIRECT LABOR	5,800	5,800	INCLUDED	121,800	INCLUDED	25,000	3,200	INCLUDED
DIRECT COSTS	24,200	24,200	8,500,000	743,500	200,000	50,000	491,900	87,400
FREIGHT, DUTY, TAXES	1,100	1,100	INCLUDED	33,200	2,400	1,700	30,600	INCLUDED
CONSTRUCTION OVERHEAD	4,400	4,400	INCLUDED	149,300	30,800	9,200	82,600	INCLUDED
ENGINEERING	1,800	1,800	INCLUDED	59,700	4,100	2,800	46,900	INCLUDED
BARE MODULE COSTS	31,500	31,500	8,500,000	985,700	247,300	63,700	652,000	87,400
CONTINGENCY	4,700	4,700	1,275,000	142,900	35,100	9,400	96,500	INCLUDED
COMPLETE MODULE COSTS	36,200	36,200	9,775,000	1,128,600	282,400	73,100	748,500	87,400
CONTRACTOR	800	800	293,300	23,700	6,000	1,600	16,500	INCLUDED
TOTAL MODULE COSTS	37,000	37,000	10,068,300	1,152,300	288,400	74,700	765,000	87,400

gives a breakdown of each of the module costs which make up the total capital cost.

Sensitivity Analysis--

Since many elements of the capital cost estimate are highly dependent on the magnitude of the input parameters, it is necessary to determine the sensitivity of this variance. This is done for the following cases:

1) Effect of fuel heating value
2) Effect of baghouse operating temperature
3) Effect of varying the sulfur content of the fuel

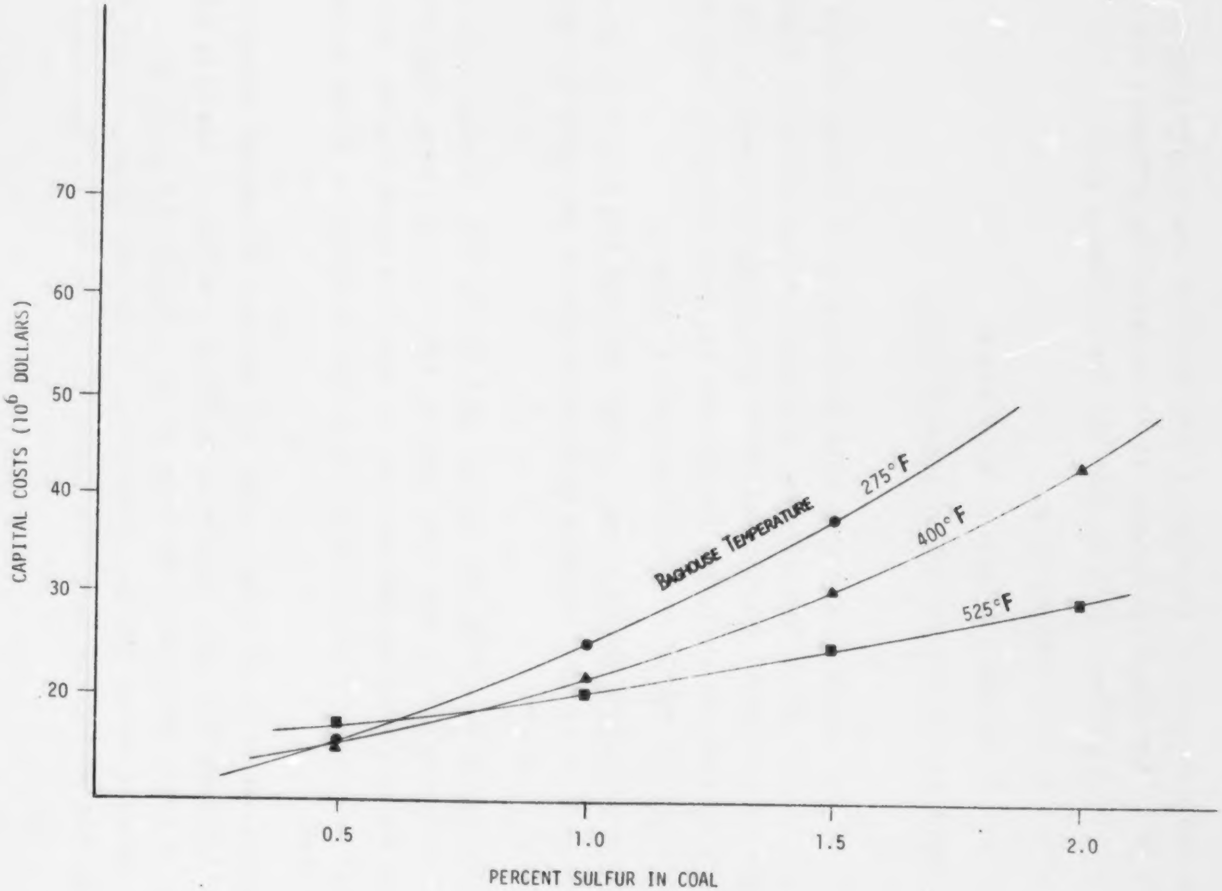
The results of this analysis are presented graphically in Figures 19 and 20. Calculated data upon which these figures are based are contained in Appendix B. In order to facilitate the discussion of the assumptions made in the capital cost sensitivity analysis, the process has been broken up into the same components that appear in Table 13, capital costs.

Nahcolite Unloading Section--The capital cost for the nahcolite unloading section was assumed to vary linearly with the weight of the nahcolite being consumed by the process.

Site Storage Silo--Since the capital cost of the site storage silo is determined by the weight of the steel used to fabricate the silo, and this, in turn, is predetermined by the quantity of nahcolite being stored, the capital cost was varied linearly with the weight of nahcolite being consumed by the process.

Conveyors--We have solicited quotes for the cost of several conveying systems (all pneumatic), each capable of handling a different quantity of nahcolite. A graph was made of the cost of the conveyor per pound of nahcolite conveyed versus the design capacity. From this graph, it was then possible to interpolate the cost of a conveying system for any load-handling capacity.

Surge Tank--As in the case of the site storage silo, the cost of the surge tank was determined by the quantity of steel necessary to manufacture



(COAL HEATING VALUE 10,500 BTU/LB)

FIGURE 19. CAPITAL COSTS VERSUS PERCENT SULFUR CONTENT IN AS-FIRED COAL FOR VARIOUS BAGHOUSE TEMPERATURES

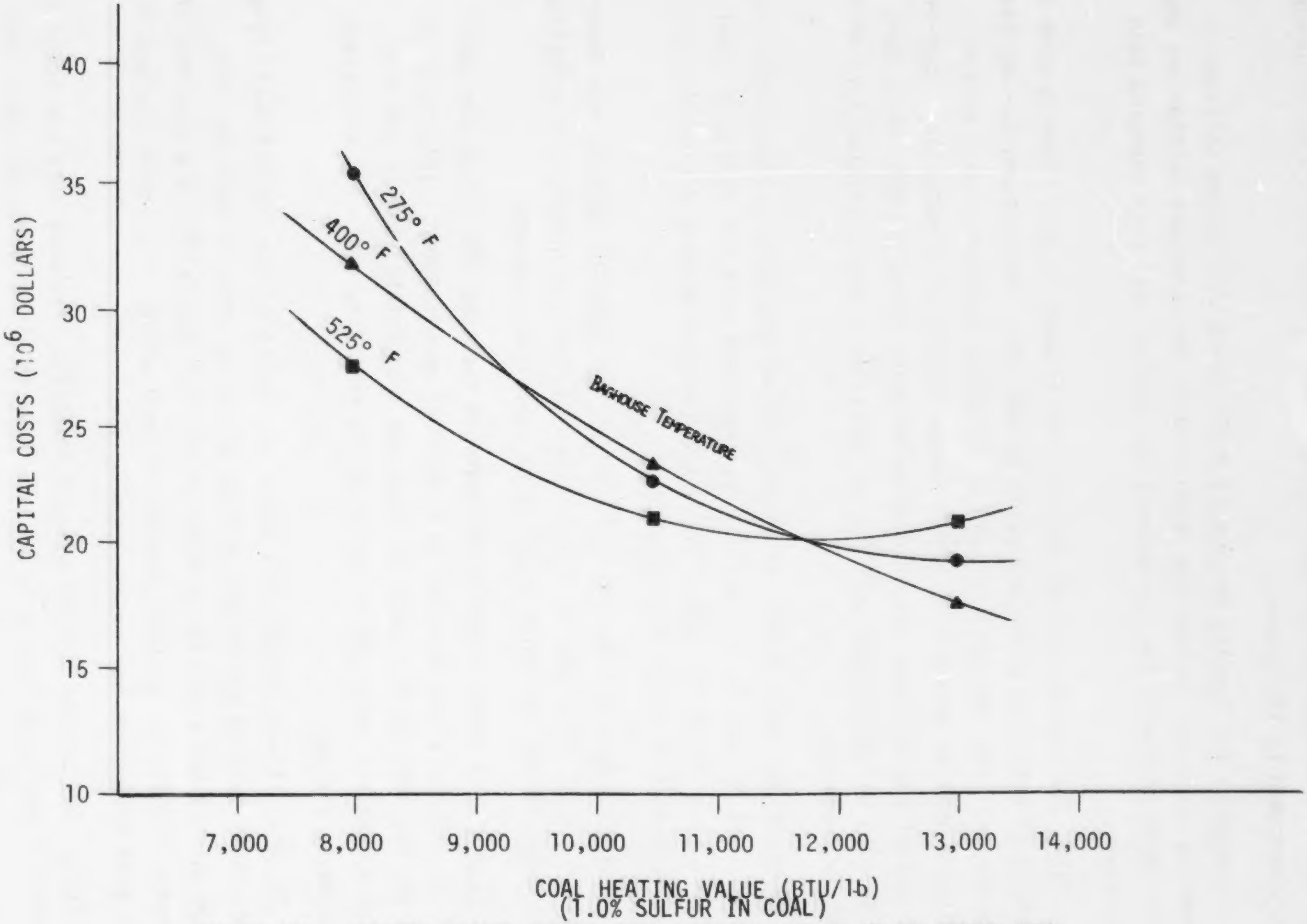


FIGURE 20. CAPITAL COSTS VERSUS COAL HEATING VALUE IN AS-FIRED COAL FOR VARIOUS BAGHOUSE TEMPERATURES

the tank. Thus, the capital cost varied linearly with the weight of nahcolite being consumed by the process.

Conveyors and Feeding Network of Mill--Since this system utilizes a pneumatic conveying system, the graph used in the conveyors section was again utilized to determine the variance of the capital cost with changing load conditions.

Mill--The capital cost of the mill was assumed to vary linearly with the quantity of nahcolite being processed by the mill. This assumption was based upon the fact that the mill is made up of module sections, i.e., several mills are used in parallel to handle large quantities of material. However, the cost of the mill was not allowed to decrease below $50,000, since this was felt to be the minimum cost for an installed product, irrespective of the quantity involved.

Storage Day Tank--Again, as in the case of the other storage tanks, the capital cost of the storage day tank is dependent upon the weight of steel utilized in erecting the tank. Thus, the cost was assumed to vary linearly with the amount of nahcolite consumed.

Injection Systems--The capital cost of the injection systems was determined by use of the graph for the variance of cost per quantity of nahcolite transported versus the total quantity of nahcolite consumed.

Baghouse--A curve relating the baghouse cost per CFM versus the total flue gas flow rate was provided by a baghouse manufacturer. This relationship was factored to represent the baghouse cost established in the base design system and then used to estimate the resulting costs for variations in baghouse sizing.

ID Fans--It was assumed that the capital cost of the induced draft fans varied linearly with the volume of flue gas being drawn through the fans. Though this is not strictly correct, it was felt that within the confines of the error bounds of the total capital cost estimation, the error involved is small with respect to the total system error.

Ducts--The capital cost of a fully installed insulated duct was found to be almost linear with flue gas flow rate down to about 1,200,000 cubic feet per minute (34,000 cubic meters per minute) and deviated about 20% (on the low side) from linearity between 400,000 cubic feet per minute (11,000 cubic meters per minute) and 1,000,000 cubic feet per minute (28,000 cubic meters per minute).

Ash Disposal Conveyor--Since a pneumatic conveyor is used for the ash disposal conveyor, the capital cost variation with changes in the quantity of waste from the baghouse was found using the graph for conveyor cost estimation. However, it was necessary to determine the capital cost of the system based on two variables--the quantity of fly ash and the quantity of spent nahcolite. Thus, because the base case input parameters dictated a baghouse waste of approximately 50% fly ash and 50% spent nahcolite, half of the base case conveyor cost is proportional to the amount of nahcolite used in the process and half the base case cost is proportional to the amount of fly ash present in the flue gas; both portions of the ash disposal conveyor cost estimate were determined utilizing the pneumatic conveyor cost estimate graph.

Ash Disposal Silo--As in the case of the ash disposal conveyor, the base case capital cost estimate for the ash disposal silo is evenly split between the cost due to the spent nahcolite and the cost due to the fly ash. However, in this case, since silo costs are directly related to the quantity of steel used in the fabrication of the silo, half the base case cost was varied linearly with nahcolite usage and half was varied linearly with fly ash generated.

Retrofit Case Evaluation--

In order to estimate the capital costs associated with retrofitting an existing pulverized coal-fired utility boiler with a nahcolite/baghouse flue gas desulfurization process, one need only consult the 275°F operating temperature lines on Figure 19 and Figure 20 showing the variance of capital costs with fuel heating value and with percent sulfur present in the fuel.

Operating Costs

The costs associated with the ongoing operation of the dry sorbent baghouse FGD system were estimated for the base design. These costs were then varied by sensitivity analysis techniques to establish the corresponding operating costs for the specific alternative operating conditions analyzed in the preceding section. Included for consideration within operating costs were pure operational costs such as raw materials, labor, and utilities consumption, plant overhead costs, and maintenance costs, including labor and materials.

Plant Availability--

In selecting the nahcolite/baghouse operating parameters, every attempt was made to keep the conditions as equivalent to the TVA study as possible. For example, in both the TVA study and the nahcolite baghouse study, a 500 megawatt, coal-fired power plant was used. Also, in both cases, the amount of SO_2 emitted per million BTU of heat input was maintained at approximately 0.5 pounds. In addition, a plant operating year of 7000 hours which gives a plant availability of 80% was assumed for both the TVA study and the nahcolite/baghouse system.

Source of Operating Costs--

Each of the various costs associated with the operation and maintenance of the dry sorbent/baghouse FGD system has been established on the basis of historical data, manufacturers recommendations, and estimations of current utility and material costs. The basis for each item has been established as follows:

Raw Materials--Since a market for nahcolite does not exist at the present time, it was only possible to estimate the cost of nahcolite. The three major lease-holders of mineral rights to land containing nahcolite deposits estimate nahcolite costs at $25 per ton for nahcolite assaying at 70% sodium bicarbonate minimum. For a more detailed discussion of nahcolite costs, see the section on sorbent supply and costs.

To estimate the shipping costs of nahcolite from the mine mouth to the proposed power plant site, it was assumed that a unit train could be utilized. It has been assumed that the average distance from any one of the proposed mine mouths to the power plant would be 750 miles (1,200 kilometers). With these assumptions, it was estimated that the cost of nahcolite transportation (to include hauling filled cars from the mine to the power plant and returning the empty cars back to the mine mouth) would be in the vicinity of one cent per ton-mile.

From the above cost estimates, it is calculated that the cost of nahcolite delivered to the power plant would be $32.50 per ton ($35.82 per metric ton).

Electrical Consumption--The only utility consumed by the nahcolite/baghouse system to any appreciable degree is electricity. To estimate the annual electrical power consumption, an inventory was made of the major consumers of electricity, such as ID fans, baghouse, conveyors and the mill. These power requirements were converted to kilowatt hours (based on a 7000 hour annual operating period) and were factored up by 10% to realize the power consumed by any ancillary equipment as well as such incidentals as lighting and environmental control. Electrical costs are estimated to be $.018/kWh.

Labor--Labor requirements were established on the basis of the complexity of the dry sorbent baghouse system, comparative requirements for similar solids handling operations, and recommendations from the baghouse manufacturers. Labor required for system maintenance and bag replacement were not included in this section but were included as part of the maintenance costs. Labor costs were based on an average of crafts and were assumed to be $8.81 per hour.

It was estimated that a total of 9.0 man-years would be needed to effectively operate the nahcolite/baghouse flue gas desulfurization process. The following breakdown is given as to the allocation of man-years for the baghouse system:

2.0 man-years to supervisory personnel

2.0 man-years to nahcolite unloading

1.0 man-years to general baghouse operation

3.0 man-years to system operations

1.0 man-years to waste unloading

Analysis--Analysis costs were estimated for the nahcolite/baghouse system by assuming the utilization of two full-time chemists at the above rate plus the additional cost of laboratory supplies.

Maintenance--Maintenance costs for the baghouse were estimated at $289,000 per year. This figure includes materials such as bag replacement every 2 years, clamps and hardware, and the labor required to perform this maintenance. This figure is based on estimates given by various baghouse manufacturers. Yearly maintenance costs for the balance of the dry sorbent system were assumed to be approximately 4 percent of the capital investment for this portion of the system.

Waste Disposal--Waste disposal costs for the nahcolite/baghouse system were estimated to be $6 per ton of combined spent nahcolite/fly ash, as outlined in the waste disposal section. A cost of $5.90 per ton has been assumed as typical landfill costs for pure fly ash. The actual waste disposal costs attributed to the nahcolite/baghouse system were determined by calculating the total landfill costs of the combined material at $6 per ton less the cost of operating a standard design landfill for the fly ash produced at $5.90 ton.

Base Case Evaluation--

An estimation of the yearly cost for operating the dry sorbent baghouse based on the above criteria has been performed. Table 14 presents these data in 1977 dollars.

Sensitivity Analysis--

For the nahcolite/baghouse flue gas desulfurization scheme, a sensitivity analysis was prepared with the following variations in operating parameters being investigated:

(1) Fuel heating value

(2) Percent sulfur in fuel

(3) Baghouse operating temperature

The results of this analysis are presented graphically in Figures 21 and 22. Calculated data upon which these figures are based are contained in Appendix C.

To simplify the discussion of the operating costs sensitivity analysis, the components of the total operating cost will be examined individually to ascertain their variance to differing input conditions.

TABLE 14. BAGHOUSE NAHCOLITE PROCESS - TOTAL AVERAGE ANNUAL OPERATING COSTS, REGULATED UTILITY ECONOMICS*

	Annual Quantity	Unit Cost, $	Total Annual Cost, $
Direct Costs			
Delivered Raw Materials			
Nahcolite	110,250 tons	32.50/ton	3,583,100
Conversion Costs			
Operating Labor and Supervision	18,000 man-hrs.	8.81/man-hr.	158,600
Utilities			
Electricity	37,620,000 kWh	0.018/kWh	677,200
Maintenance			
Labor and Materials			713,600
Analyses			55,700
SUBTOTAL DIRECT COSTS.			5,188,200
Indirect Costs			
Average Capital Charges at 14.9% of Total Capital Investment			3,081,700
Overhead			
Plant, 20% of Conversion Costs			167,200
Administrative, 10% of Operating Labor			15,900
SUBTOTAL INDIRECT COSTS.			3,264,800
Annual Cost for Trucking and Off-Site Disposal of Sodium Solids at $6.00/ton			496,100
TOTAL ANNUAL OPERATING COSTS			8,949,100

*500MW new coal-fired power plant, 1.0% S in fuel; 70% SO_2 removal; off-site solids disposal.

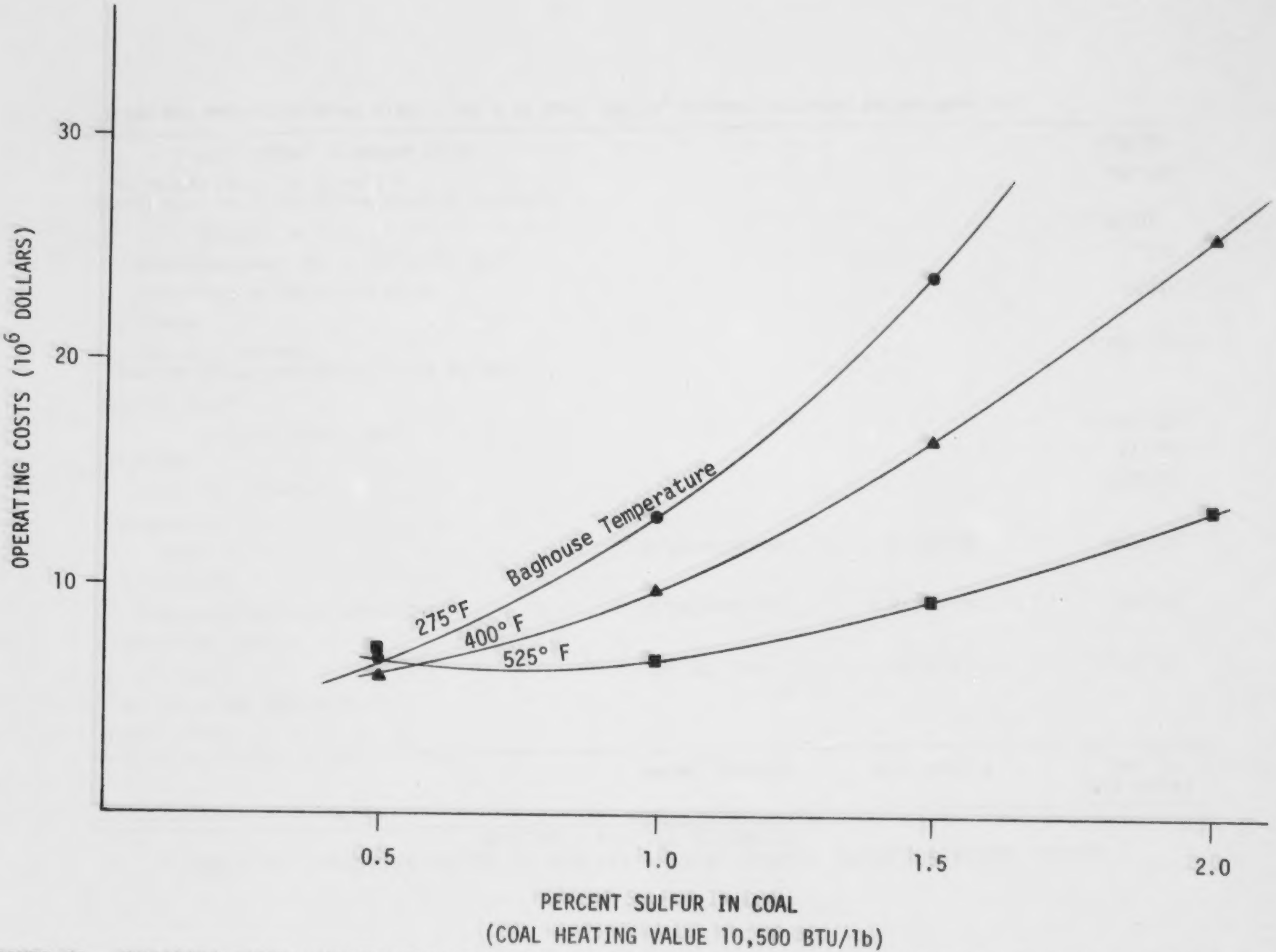


FIGURE 21. OPERATING COSTS VERSUS PERCENT SULFUR CONTENT IN AS-FIRED COAL FOR VARIOUS BAGHOUSE TEMPERATURES

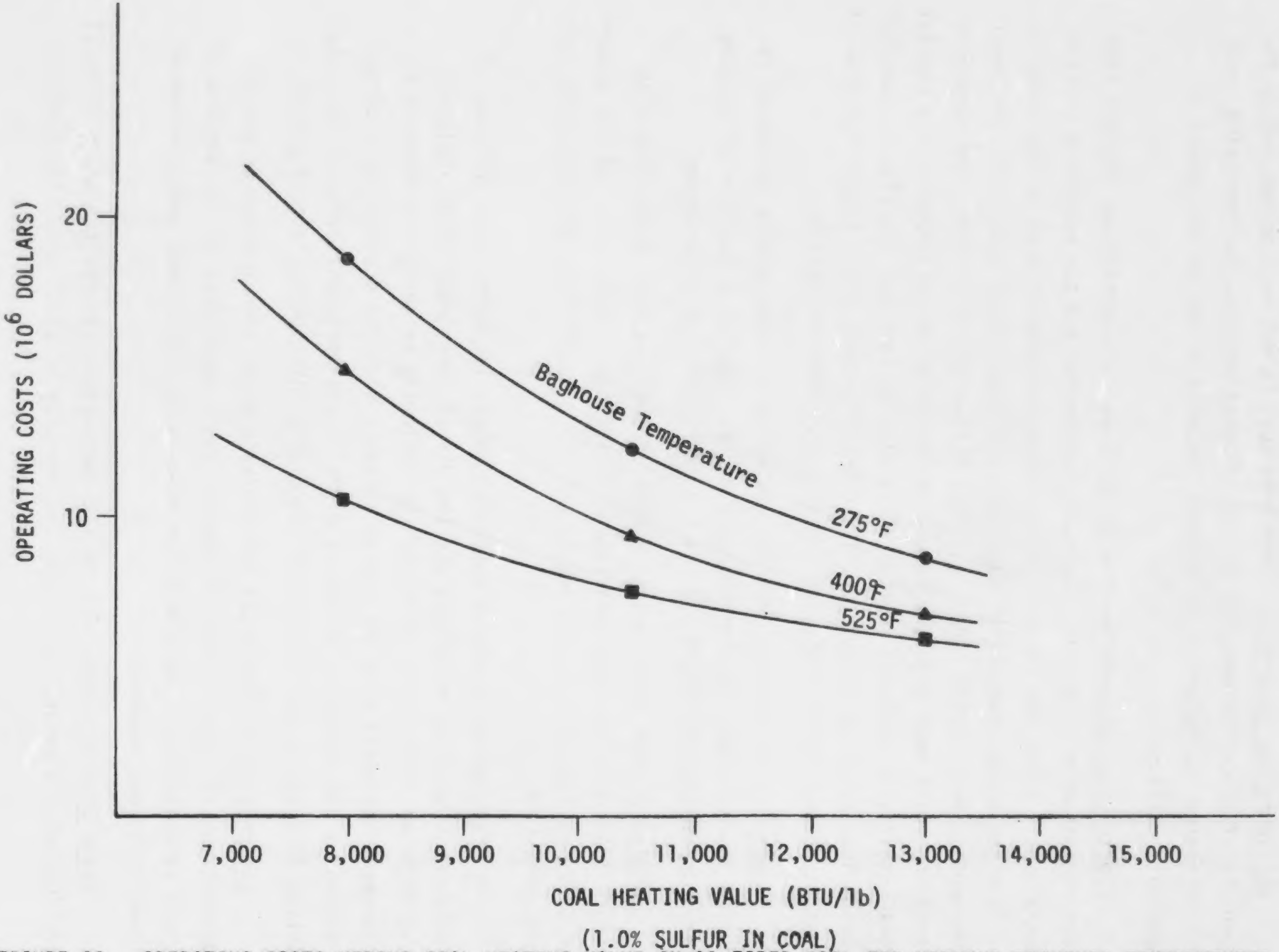


FIGURE 22. OPERATING COSTS VERSUS COAL HEATING VALUE IN AS-FIRED COAL FOR VARIOUS BAGHOUSE TEMPERATURES

Delivered Raw Materials--Since this cost is entirely determined by the quantity of nahcolite consumed in the proposed process, the operating cost for delivered raw materials is linearly related to the annual amount of nahcolite utilized.

Electricity Consumption--Electrical power consumption was divided into four constituents: ID fans, baghouse, conveyors, and the nahcolite milling operations. Since the ID fans and baghouse are proportional to the quantity of flue gas being transported through the system, their portion of the power consumption were varied linearly with the flue gas flow rate. The conveyors operating costs were adjusted according to the graph developed in the capital cost sensitivity analysis section which plotted cost per quantity transported versus load handling capacity. Mill operating costs are a linear function of quantity of nahcolite processed and were adjusted accordingly.

Labor--It was assumed that the operating and supervisory personnel required to adequately run the baghouse process was not a function of varying input parameters and therefore was held constant at 9.0 man-years.

Analyses--Because a certain number of analyses must be performed to ensure the proper working environment in the baghouse process and the number of analyses done is not affected by the input conditions, the operating costs were held constant.

Maintenance--Maintenance operating costs, including both the cost of replacement baghouse materials and the direct investment costs excluding the baghouse costs, were varied in the following manner: the baghouse replacement materials cost was varied linearly with the quantity of flue gas handled by the baghouse, while the direct investment cost was a direct computation from the direct investment associated with the applicable process

Indirect Costs--Since all the indirect costs are computed as certain percentages of costs already discussed, their dependence on the varying of input parameters will vary with the base costs. No direct manipulation of these costs was required.

Waste Disposal Costs--Due to the operating costs for the waste disposal section being dependent only upon the quantity of spent nahcolite (soluble sodium salts) being produced by the baghouse system, the costs were varied linearly with the quantity of nahcolite consumed in the process.

Retrofit Case Evaluation--

As in the case of the capital cost retrofit case evaluation, an estimate of the operating costs for the nahcolite/baghouse flue gas desulfurization process can be obtained by using the 275°F operating temperature line for the plots depicting operating costs versus percent sulfur in coal and operating costs versus coal heating value. The 275°F operating temperature corresponds to the typical air preheater exit temperature of most existing utility boilers of comparable size.

Comparison with Limestone Scrubber--

In order to fully evaluate the cost impact of the nahcolite/baghouse process, it was necessary to develop accurate cost projections for alternative flue gas desulfurization processes. This analysis has been limited to consideration of the limestone scrubber, as this is the prominant method available today and also presents the least-cost option of the presently available systems. The cost projections for a limestone flue gas desulfurization scrubbing system attached to a new 500 MW coal-fired power plant, located in the area of study, were developed from a TVA computer program. Input criteria for this program were chosen in such a way that the calculated flue gas streams from the boiler matched the assumed flue gas stream composition for the nahcolite/baghouse system as described elsewhere in this report.

Since the nahcolite/baghouse process accomplishes both fly ash particle removal and flue gas desulfurization within the same piece of equipment, additional costs have been included within the limestone scrubber cost projections to include the cost of comparable particle removal. This particle removal is equivalent to the Federal New Source Performance Standards of 0.1 lb particulate matter per million Btu heat input, and requires a particle removal rate of 98.8% of all entrained fly ash. The TVA program allowed consideration of two configurations for particle control. The first of these involved no upstream particle removal but took credit for 40% particle removal within the limestone scrubber itself. The second alternative included a 33% efficient mechanical collector upstream of the scrubber. Since this second alternative would still require an electrostatic precipitator or equivalent device capable of removing 97.3% of the particulate matter in order to guarantee achievement of the New Source Performance Standard at the outlet of the scrubber, the cost of this alternative was demonstratively higher than the first alternative, and hence, was not included in the detailed cost comparison.

Cost projections for electrostatic precipitator of equivalent performance for the nahcolite/baghouse system were determined from projections made by the Industrial Environmental Research Laboratory of the Environmental Protection Agency. These projections indicated that a 500 MW power plant burning 1% sulfur coal with a 10% ash content would require a total turn key cost for the electrostatic precipitator of 28.5 million dollars (1977 dollars). Operating and maintenance costs were further determined from this program to be $250,000 per year. The annualized operating cost for trucking and off-site disposal of the fly ash constituent of the collected particulate matter was not included in the dry sorbent/baghouse FGD system. This cost has therefore not been included within the annualized operating costs for the limestone scrubber system. Table 15 presents a comparison of the capital costs of the dry sorbent/FGD baghouse system and the limestone slurry FGD system.

Annualized operating costs for the limestone scrubber FGD system have been established by using the TVA cost projections program in conjunction with site specific cost data developed for the western nahcolite baghouse system. Additionally, sources of limestone in the Green River, Wyoming area were contacted to determine the realistic cost of obtaining limestone at the plant site in the quantities needed for the equivalent limestone slurry process. Table 16 presents a breakdown of the total average annual operating costs for the limestone scrubber system assumed for this cost comparison. A comparison of the annualized operating costs for the nahcolite/baghouse FGD system and the limestone slurry FGD system is presented in Table 17.

TABLE 15. COST COMPARISON OF DRY SORBENT/BAGHOUSE FGD SYSTEM WITH LIMESTONE SLURRY FGD PROCESS

	IN 1977 DOLLARS (10^3 DOLLARS)	
	BAGHOUSE NAHCOLITE PROCESS (400°F)	LIMESTONE SLURRY PROCESS
EQUIPMENT		
M[1]	(COMBINED)	12099
L[1]	16727	4072
MATERIALS HANDLING		
M[1]	191	680
L[1]	85	293
PARTICLE REMOVAL	---	28500
EXCAVATION	2	1386
FOUNDATIONS		
M	24	123
L[2]	96	510
ENGINEERING	547*	1803
CONTRACTOR	518	681
CONTINGENCY	2596	4657
OTHER CONSTRUCTION COSTS	1969	4476
TOTAL	22755	59280

M = MATERIALS, L = LABOR

[1] MINIMUM IN-PROCESS STORAGE; ONLY PUMPS ARE SPARED.

[2] CONSTRUCTION LABOR SHORTAGES WITH ACCOMPANYING OVERTIME PAY INCENTIVE NOT CONSIDERED.

* ENGINEERING COSTS FOR THE BASIC BAGHOUSE PACKAGE IS INCLUDED IN EQUIPMENT COSTS.

TABLE 16. LIMESTONE SCRUBBER - TOTAL AVERAGE ANNUAL OPERATING COSTS, REGULATED UTILITY ECONOMICS

	Annual Quantity	Unit Cost, $	Total Annual Cost, $
Direct Costs			
Raw Materials			
Limestone	35,000 tons	$10.32/ton	361,200
Subtotal Raw Materials			
Conversion Costs			
Operating Labor and Supervision	20,700 man-hr.	8.81/man-hr.	182,400
Utilities			
Steam	644,310 M lbs.	1.26/M lbs.	811,800
Process Water	205,960 M gal.	0.0951/M gal.	19,600
Electricity	49,090,180 kWh	0.018/kWh	883,600
Maintenance			
Labor and Material			1,891,100
Analyses	2,990 hr.	15.00/hr.	44,900
Operation and Maintenance of ESP			250,000
SUBTOTAL DIRECT COSTS.			4,444,600
Indirect Costs			
Average Capital Charges at 14.9% of Total Capital Investment			8,832,700
Overhead			
Plant, 20% of Conversion Costs			816,700
Administrative, 10% of Operating Labor			18,200
SUBTOTAL INDIRECT COSTS.			9,667,500
Annual Cost for Trucking and Off-Site Disposal of Calcium Solids at $4.364/Ton			183,300
TOTAL ANNUAL OPERATING COSTS			**14,295,500**

TABLE 17. COMPARISON OF TOTAL AVERAGE ANNUAL OPERATING COSTS FLUE GAS DESULFURIZATION PROCESSES

	NAHCOLITE/ BAGHOUSE PROCESS (400°F)	LIMESTONE SLURRY PROCESS
	(IN 1977 DOLLARS)	
DIRECT COSTS	5,188,200	4,444,600
INDIRECT COSTS	3,264,800	9,667,600
WASTE DISPOSAL	496,100	183,300
TOTAL	8,949,100	14,295,500

SECTION 6

ENGLISH TO METRIC CONVERSION

CONVERTING UNITS OF MEASURE

EPA policy is to express all measurements in metric units. When implementing this practice will result in undue cost or lack of clarity, conversion factors are provided for the nonmetric units used in this report. Table 18 below provides appropriate factors for this report.

TABLE 18. CONVERSION FACTORS

Multiply English Unit	By Conversion	To Obtain Metric Unit
Acre-Feet	1233.5	Cubic Meters
British Thermal Unit	1054.8	Joules
British Thermal Unit/Pound	2325.4	Joules/Kilogram
Cubic Feet/Minute	0.028	Cubic Meters/Minute
Cubic Feet	0.028	Cubic Meters
Cubic Feet	28.32	Liters
Degree Fahrenheit	0.555(°F-32)	Degree Centigrade
Feet	0.3048	Meters
Gallon	3.785	Liters
Gallon/Minute	0.0631	Liters/Second
Grains	0.06480	Grams
Grains/ft^3	2.288	Grams/m^3
Inches	2.54	Centimeters
Pounds	0.454	Kilograms
Mile	1.609	Kilometer
Pound-Moles	453.6	Gram-Moles
Tons (Short)	0.907	Metric Tons (1000 Kilograms)

SECTION 7

REFERENCES

1. TRW. Evaluation of Dry Sorbent and Fabric Filtration for FGD. EPA Contract No. 68-02-1412, Task Order No. 19. 28 June 1976.

2. McGlamery, G. G., R. L. Torstrick, W. J. Broadfoot, J. P. Simpson, L. J. Henson, S. V. Tomlinson, and J. F. Young. Detailed Cost Estimates for Advanced Effluent Desulfurization Processes. EPA-600/2-75-006, Office of Research and Development, U.S. Environmental Protection Agency. January 1975. 418 pp.

3. Bradway, R. M., and R. W. Cass. Fractional Efficiency of a Utility Boiler Baghouse NUCLA Generating Plant. EPA-600/2-75-013-a, Office of Research and Development, U.S. Environmental Protection Agency. August 1975. 137 pp.

4. Cass, R. W., and R. M. Bradway. Fractional Efficiency of a Utility Boiler Baghouse: Sunbury Steam-Electric Station. Office of Research and Development, U.S. Environmental Protection Agency. March 1976. 231 pp.

5. Rittenhouse, R. C. New Generating Capacity: Who's Doing What. Power Engineering. Volume 81. Number Four. April 1977. pp. 62-69.

6. Potter, P. J. Power Plant Theory and Design. The Ronald Press Company, New York, New York, Second Edition, 1959. 702 pp.

7. Adams, R. C., T. E. Eggleston, J. L. Haslbeck, R. C. Jordan, and E. Pulaski. Demonstration of Wellman-Lord/Allied Chemical FGD Technology: Boiler Operating Characteristics. EPA-600/7-77-014, Office of Research and Development, U.S. Environmental Protection Agency. February 1977. 254 pp.

8. Bechtel Corporation. Evaluation of Dry Alkalis for Removing Sulfur Dioxide from Boiler Flue Gases. EPRI FP-207 (Research Project 491-1). Final Report. October 1976. 144 pp.

9. Liu, H., and R. Chaffee. Final Report on Evaluation of Fabric Filter as Chemical Contactor for Control of Sulfur Dioxide from Flue Gas. Air Preheater Company, Incorporated. December 1969. 159 pp.

10. Doyle, D. J. Fabric and Additive Remove SO_2. Electrical World. February 15, 1977. pp. 32-34.

11. Genco, J. M. and H. S. Rosenberg. Sorption of SO_2 on Ground Nahcolite Ore. Journal of the Air Pollution Control Association. Volume 26. Number 10. October 1976. pp. 989-990.

12. Dulin, J. M. and E. C. Rosar. Advances in Disposal of Sodium Salts Wastes Produced by SO_x Emissions Control Processes, Cooling Towers, and Water Demineralizers. Presented at the Annual Meeting of the Air Pollution Control Association, Session No. 9, "Waste Disposal Without Environmental Degradation." June 11, 1974. 31 pp.

13. Proceedings: Symposium on Flue Gas Desulfurization. New Orleans, Louisiana. March 1976. Volume II. EPA-600/2-76-136b, Office of Research and Development, U.S. Environmental Protection Agency. May 1976. pp. 565-700.

14. Proceedings: Symposium on Flue Gas Desulfurization - Atlanta, November 1974. Volume II. EPA-650/2-74-126b, Office of Research and Development, U.S. Environmental Protection Agency. December 1974. pp. 851-1028.

15. Jones, J. W. Disposal of Flue-Gas-Cleaning Wastes. Chemical Engineering Volume 84. Number 4. February 14, 1977. pp. 79-85.

16. Burnham, A. K., G. V. Clader, J. S. Fritz, G. A. Junk, H. J. Svec, and R. Vick. Identification and Estimation of Neutral Organic Contaminants in Potable Water. Analytical Chemistry. Volume 44. 1972. pp. 139.

17. Keith, L. H., and S. H. Hercules. Environmental Applications of Advanced Instrumental Analyses: Assistance Projects, FY 69-71. EPA-R2-73-155, Office of Research and Monitoring, U.S. Environmental Protection Agency. May 1973. 83 pp.

18. Ghassemi, M. Analysis of a Land Disposal Damage Incident Involving Hazardous Waste Materials: Dover Township, New Jersey. EPA Contract No. 68-01-2956, Task Order 68-01-3187, Office of Solid Waste Management Programs, U.S. Environmental Protection Agency. May 1976. 121 pp.

19. Sorg, T. J., and H. L. Hickman, Jr. Sanitary Landfill Facts. Public Health Service Publication No. 1792. Reprinted, EPA, 1971. 30 pp.

20. Guthrie, K. M. Capital Cost Estimating. Chemical Engineering. Volume 76. Number 6. March 24, 1969. pp. 114-142.

21. Personal Correspondence with Mr. J. Phelan, Wheelabrator-Frye 2/77.

22. Personal Correspondence with Mr. J. H. Smith, Rock School Corp. 1977.

23. Personal Correspondence with Mr. Jacques Dulin, Industrial Resources Inc., 1977.

APPENDIX A

SAMPLE CALCULATIONS FOR DETERMINATION OF CAPITAL COSTS FROM BASIC MATRIX

(All subscripts refer to year for which parameters are valid)

E_{1977} = equipment costs = 495000

$$E_{1968} = \frac{E_{1977}}{\text{inflator* for pumps and motors}}$$

(*inflators derived in Table 11)

$$E_{1968} = \frac{495000}{2.0321} = 243591.54$$

$$\begin{aligned} m_{1968} &= \text{field material lost costs} = (k_{1,\ 1968})(E_{1968}) \\ &= (0.27)(243591.54) \\ &= 65769.715 \end{aligned}$$

$$\begin{aligned} m_{1977} &= (\text{inflator for general equipment costs})(m_{1968}) \\ &= (1.9270)(65769.715) \\ &= 126735.68 \end{aligned}$$

$$\rightarrow k_{1,\ 1977} = \frac{m_{1977}}{E_{1977}} = \frac{126735.68}{495000}$$

$$k_{1,\ 1977} = 0.256$$

L_{1968} = direct labor costs = $(k_{2, 1968})(E_{1968})$

= (0.34)(243591.54)

= 82821.123

L_{1977} = (inflator for labor rates)(L_{1968})

= (1.4699)(82821.123)

= 121742.89

$$\rightarrow k_{2, 1977} = \frac{L_{1977}}{E_{1977}} = \frac{121742.89}{495000}$$

$$k_{2, 1977} = 0.246$$

f_{1968} = freight, duty, and taxes costs = $(k_{3, 1968})(E_{1968})$

= (0.08) (243591.54)

= 19487.323

f_{1977} = (inflator for freight, duty, and taxes)(f_{1968})

= (1.7030)(19487.323)

= 33186.911

$$\rightarrow k_{3, 1977} = \frac{f_{1977}}{E_{1977}} = \frac{33186.911}{495000}$$

$$k_{3, 1977} = 0.067$$

M_{1968} = direct material costs = $E_{1968} + m_{1968}$

$= 243591.54 + 65769.715$

$= 309361.255$

$M_{1968} + L_{1968}$ = direct costs

$= 309361.255 + 82821.123$

$= 392182.378$

c_{1968} = construction overhead costs = $(k_{4,\ 1968})(M_{1968} + L_{1968})$

$= (0.178)(392182.378)$

$= 69808.461$

c_{1977} = (inflator for construction overhead)(c_{1968})

$= (1.8821)(69808.461)$

$= 131386.5$

$(M_{1977}) + L_{1977} = (E_{1977} + m_{1977}) + L_{1977}$

$= (495000 + 126735.68) + 121742.89$

$= 743478.57$

$$\rightarrow k_{4,\ 1977} = \frac{c_{1977}}{M_{1977} + L_{1977}} = \frac{131386.5}{743478.57}$$

$$k_{4},\ 1977 = 0.177$$

e_{1968} = engineering costs = $(k_{5,\ 1968})(M_{1968})$

= (0.136)(309361.25)

= 42073.13

e_{1977} = (inflator for engineering rates)(e_{1968})

= (1.4152)(42073.13)

= 59542.54

$$\rightarrow k_{5,\ 1977} = \frac{e_{1977}}{M_{1977}} = \frac{59542.54}{621735.68}$$

$$k_{5,\ 1977} = 0.096$$

B_{1968} = bare module costs = $(M_{1968} + L_{1968}) + f_{1968} + c_{1968} + e_{1968}$

= (392182.378) + 19487.323 + 69808.461 + 42073.13

= 523551.292

h_{1968} = contingency costs = $(k_{6,\ 1968})(B_{1968})$

= (0.15)(523551.292)

= 78523.693

h_{1977} = (inflator for contingency)(h_{1968})

= (1.7841)(78523.693)

= 140099.41

$B_{1977} = (M_{1977} + L_{1977}) + f_{1977} + c_{1977} + e_{1977}$

= (743478.57) + 33186.911 + 131386.5 + 59542.54

= 967594.521

$$\rightarrow k_{6,\ 1977} = \frac{h_{1977}}{B_{1977}} = \frac{140099.41}{967594.521}$$

$$k_{6,\ 1977} = 0.145$$

S_{1968} = complete module costs = $B_{1968} + h_{1968}$

$= 523551.292 + 78523.693$

$= 602083.985$

F_{1968} = contractor costs = $(k_{7,\ 1968})(S_{1968})$

$= (0.026)(602083.985)$

$= 15654.183$

F_{1977} = (inflator for contractor)(15654.183)

$= (1.5023)(15654.183)$

$= 23516.851$

$S_{1977} = (M_{1977} + L_{1977}) + f_{1977} + c_{1977} + e_{1977} = h_{1977}$

$= 743478.57 + 33186.911 + 131386.5 + 59542.54 + 140099.41$

$= 1107693.9$

$$\rightarrow k_{7,\ 1977} = \frac{F_{1977}}{S_{1977}} = \frac{23516.851}{1107693.9}$$

$$k_{7,\ 1977} = 0.021$$

APPENDIX B

CAPITAL COSTS FOR VARYING INPUT PARAMETERS

CAPITAL COST SENSITIVITY ANALYSIS

COAL ANALYSIS: % S = 1.0 HEATING VALUE = 8,000

ITEM	275°F	400°F	525°F
A. DESIGN PARAMETERS			
Flue Gas Flow	1.68×10^6	1.97×10^6	2.25×10^6
Moles SO_2/hr	168	168	168
Efficiency	.79	.77	.79
Nahcolite Usage (lb/hr)	69,200	52,600	31,900
B. ITEM COSTS			
Unloading	12,340,700	9,380,300	5,353,900
Site Storage	4,667,800	3,548,000	2,151,800
Conveyor	369,200	280,600	178,700
Surge Tank	194,700	148,000	89,700
Conveyor	299,700	227,800	145,100
Mill	966,500	734,600	445,500
Storage Tank	170,200	129,400	78,500
Injection Conveyor	80,100	60,900	38,800
Injection Conveyor	80,100	60,900	38,800
Baghouse	10,792,300	11,591,100	12,812,000
I.D. Fans	1,290,600	1,513,400	1,728,500
Ducts	305,700	330,200	351,400
Ash Conveyor	118,200	98,800	76,500
Ash Silo	1,281,300	1,066,900	799,500
Air Preheater	2,186,200	2,563,600	2,928,000
Waste Disposal	87,400	87,400	87,400
TOTAL	35,230,700	31,821,900	27,304,100

CAPITAL COST SENSITIVITY ANALYSIS

COAL ANALYSIS: % S = 1.0 HEATING VALUE = 10,500

ITEM	275°F	400°F	525°F
A. DESIGN PARAMETERS			
Flue Gas Flow	1.28×10^6	1.5×10^6	1.72×10^6
Moles SO_2/hr	128	128	128
Efficiency	.70	.70	.70
Nahcolite Usage (lb/hr)	44,800	30,000	21,500
B. ITEM COSTS			
Unloading	7,989,300	5,350,000	3,834,200
Site Storage	3,021,900	2,023,600	1,450,200
Conveyor	239,000	170,600	131,900
Surge Tank	126,100	84,400	60,500
Conveyor	194,000	138,500	107,100
Mill	625,700	419,000	300,300
Storage Tank	110,200	73,800	52,900
Injection Conveyor	51,800	37,000	28,600
Injection Conveyor	51,800	37,000	28,600
Baghouse	9,157,100	10,068,300	10,798,000
I.D. Fans	983,300	1,152,300	1,321,300
Ducts	226,900	288,400	309,200
Ash Conveyor	89,700	74,700	66,200
Ash Silo	966,200	765,000	665,200
Air Preheater	1,665,700	1,952,000	2,238,300
Waste Disposal	87,400	87,400	87,400
TOTAL	25,586,100	22,722,000	21,479,900

CAPITAL COST SENSITIVITY ANALYSIS

COAL ANALYSIS: % S = 1.0 HEATING VALUE = 13,000

ITEM	275°F	400°F	525°F
A. DESIGN PARAMETERS			
Flue Gas Flow	1.04×10^6	1.21×10^6	1.39×10^6
Moles SO_2/hr	103	103	103
Efficiency	.63	.63	.63
Nahcolite Usage (lb/hr)	29,400	19,700	15,600
B. ITEM COSTS			
Unloading	5,243,000	3,513,200	2,783,000
Site Storage	1,983,100	1,328,800	1,052,300
Conveyor	169,400	124,000	108,200
Surge Tank	82,700	54,400	43,900
Conveyor	137,500	100,700	87,800
Mill	410,600	275,100	217,900
Storage Tank	72,300	48,500	38,400
Injection Conveyor	36,700	26,900	23,500
Injection Conveyor	36,700	26,900	23,500
Baghouse	8,047,500	8,833,000	9,538,000
I.D. Fans	798,900	929,500	1,067,800
Ducts	240,500	259,600	278,000
Ash Conveyor	74,400	64,500	61,000
Ash Silo	767,300	642,000	589,000
Air Preheater	1,353,400	1,576,600	1,808,900
Waste Disposal	87,400	87,400	87,400
TOTAL	19,541,400	17,891,100	17,808,600

CAPITAL COST SENSITIVITY ANALYSIS

COAL ANALYSIS: % S = 0.5 HEATING VALUE = 10,500

ITEM	275°F	400°F	525°F
A. DESIGN PARAMETERS			
Flue Gas Flow	1.28×10^6	1.50×10^6	1.72×10^6
Moles SO_2/hr	64	64	64
Efficiency	.40	.40	.40
Nahcolite Usage (lb/hr)	10,400	6,100	6,100
B. ITEM COSTS			
Unloading	1,854,700	1,087,800	1,087,800
Site Storage	701,500	411,500	411,500
Conveyor	96,500	71,600	71,600
Surge Tank	29,300	17,200	17,200
Conveyor	78,400	58,100	58,100
Mill	145,300	85,200	85,200
Storage Tank	25,600	15,000	15,000
Injection Conveyor	20,900	15,500	15,500
Injection Conveyor	20,900	15,500	15,500
Baghouse	9,157,100	10,068,300	10,798,200
I.D. Fans	983,300	1,152,300	1,321,300
Ducts	266,900	288,400	309,200
Ash Conveyor	58,500	53,000	53,000
Ash Silo	521,800	466,300	466,300
Air Preheater	1,665,700	1,952,000	2,238,300
Waste Disposal	87,400	87,400	87,400
TOTAL	15,713,800	15,845,100	17,051,100

CAPITAL COST SENSITIVITY ANALYSIS

COAL ANALYSIS: % S = 1.5 HEATING VALUE = 10,500

ITEM	275°F	400°F	525°F
A. DESIGN PARAMETERS			
Flue Gas Flow	1.28×10^6	1.50×10^6	1.72×10^6
Moles SO_2/hr	191	191	191
Efficiency	.80	.80	.80
Nahcolite Usage (lb/hr)	89,400	64,300	36,700
B. ITEM COSTS			
Unloading	15,943,000	11,466,800	6,544,800
Site Storage	6,030,300	4,337,200	2,475,500
Conveyor	476,900	343,000	202,000
Surge Tank	251,500	180,900	103,200
Conveyor	387,200	278,500	164,000
Mill	1,248,600	898,100	512,600
Storage Tank	219,900	158,200	90,300
Injection Conveyor	103,400	74,400	43,800
Injection Conveyor	103,400	74,400	43,800
Baghouse	9,157,100	10,068,300	10,798,200
I.D. Fans	983,300	1,152,300	1,321,300
Ducts	266,900	288,400	309,200
Ash Conveyor	141,800	112,400	81,600
Ash Silo	1,542,300	1,218,000	861,500
Air Preheater	1,665,700	1,952,000	2,238,300
Waste Disposal	87,400	87,400	87,400
TOTAL	38,608,700	32,690,300	25,877,500

CAPITAL COST SENSITIVITY ANALYSIS

COAL ANALYSIS: % S = 2.0 HEATING VALUE = 10,500

ITEM	400°F	525°F
A. DESIGN PARAMETERS		
Flue Gas Flow	1.50×10^6	1.72×10^6
Moles SO_2/hr	255	255
Efficiency	.85	.85
Nahcolite Usage (lb/hr)	100,000	52,000
B. ITEM COSTS		
Unloading	17,833,300	9,273,300
Site Storage	6,745,300	3,507,600
Conveyor	533,500	277,400
Surge Tank	281,300	146,300
Conveyor	433,100	225,200
Mill	1,396,700	726,300
Storage Tank	246,000	127,900
Injection Conveyor	115,700	60,200
Injection Conveyor	115,700	60,200
Baghouse	10,068,300	10,798,200
I.D. Fans	1,152,300	1,321,300
Ducts	288,400	309,200
Ash Conveyor	154,100	98,100
Ash Silo	1,679,200	1,059,200
Air Preheater	1,952,000	2,238,300
Waste Disposal	87,400	87,400
TOTAL	43,082,300	30,316,100

APPENDIX C

OPERATING COSTS FOR VARYING INPUT PARAMETERS

COAL ANALYSIS: % S = 0.5 HEATING VALUE = 10,500 TEMPERATURE = 275°F

	Annual Quantity	Unit Cost, $	Total Annual Cost, $
Direct Costs			
Delivered Raw Material			
Nahcolite	36,400 tons	32.50/ ton	1,183,000
Conversion Costs			
Operating Labor and Supervision	18,000/ man-hr.	8.81/ man-hr.	158,600
Utilities			
Electricity	29,467,379	0.018/ kWh	530,400
Maintenance			
Labor and Materials			608,900
Analyses			49,400
SUBTOTAL DIRECT COSTS.			2,530,300
Indirect Costs			
Average Capital Charges at 14.9% of Total Capital Investment			2,341,400
Overhead			
Plant, 20% of Conversion Costs			269,500
Administrative, 10% of Operating Labor			15,900
SUBTOTAL INDIRECT COSTS.			2,626,800
Annual Cost for Sodium Disposal at $6.00/ton			163,800
TOTAL ANNUAL OPERATING COSTS			5,320,900

COAL ANALYSIS: % S = 1.0 HEATING VALUE = 10,500 TEMPERATURE = 275°F

	Annual Quantity	Unit Cost, $	Total Annual Cost, $
Direct Costs			
Delivered Raw Material			
Nahcolite	156,800 tons	32.50/ ton	5,096,000
Conversion Costs			
Operating Labor and Supervision	18,000 / man-hr.	8.81/ man-hr.	158,600
Utilities			
Electricity	36,265,777	0.018/ kWh	652,800
Maintenance			
Labor and Materials			608,900
Analyses			49,400
SUBTOTAL DIRECT COSTS.			6,565,700
Indirect Costs			
Average Capital Charges at 14.9% of Total Capital Investment			3,812,300
Overhead			
Plant, 20% of Conversion Costs			293,900
Administrative, 10% of Operating Labor			15,900
SUBTOTAL INDIRECT COSTS.			4,122,100
Annual Cost for Sodium Disposal at $6.00/ton			705,600
TOTAL ANNUAL OPERATING COSTS			11,393,400

COAL ANALYSIS: % S = 1.5 HEATING VALUE = 10,500 TEMPERATURE = 275°F

	Annual Quantity	Unit Cost, $	Total Annual Cost, $
Direct Costs			
Delivered Raw Material			
Nahcolite	312,900 tons	32.50/ ton	10,169,300
Conversion Costs			
Operating Labor and Supervision	18,000/ man-hr.	8.81/ man-hr.	158,600
Utilities			
Electricity	46,509,309	0.018/ kWh	837,200
Maintenance			
Labor and Materials			608,900
Analyses			49,400
SUBTOTAL DIRECT COSTS.			11,823,400
Indirect Costs			
Average Capital Charges at 14.9% of Total Capital Investment			5,752,700
Overhead			
Plant, 20% of Conversion Costs			330,800
Administrative, 10% of Operating Labor			15,900
SUBTOTAL INDIRECT COSTS.			6,099,400
Annual Cost for Sodium Disposal at $6.00/ton			1,408,100
TOTAL ANNUAL OPERATING COSTS			19,330,900

COAL ANALYSIS: % S = 0.5 HEATING VALUE = 10,500 TEMPERATURE = 400°F

	Annual Quantity	Unit Cost, $	Total Annual Cost, $
Direct Costs			
Delivered Raw Material			
Nahcolite	21,350 tons	32.50/ ton	693,900
Conversion Costs			
Operating Labor and Supervision	18,000/ man-hr.	8.81/ man-hr.	158,600
Utilities			
Electricity	32,883,803	0.018/ kWh	591,900
Maintenance			
Labor and Materials			713,600
Analyses			49,400
SUBTOTAL DIRECT COSTS.			2,207,400
Indirect Costs			
Average Capital Charges at 14.9% of Total Capital Investment			2,360,900
Overhead			
Plant, 20% of Conversion Costs			302,700
Administrative, 10% of Operating Labor			15,900
SUBTOTAL INDIRECT COSTS.			2,679,500
Annual Cost for Sodium Disposal at $6.00/ton			96,100
TOTAL ANNUAL OPERATING COSTS			4,983,000

COAL ANALYSIS: % S = 1.0 HEATING VALUE = 10,500 TEMPERATURE = 400°F

	Annual Quantity	Unit Cost, $	Total Annual Cost, $
Direct Costs			
Delivered Raw Material			
Nahcolite	110,250 tons	32.50/ ton	3,583,100
Conversion Costs			
Operating Labor and Supervision	18,000/ man-hr.	8.81/ man-hr.	158,600
Utilities			
Electricity	37,620,000	0.018/ kWh	677,200
Maintenance			
Labor and Materials			713,600
Analyses			55,700
SUBTOTAL DIRECT COSTS			5,188,200
Indirect Costs			
Average Capital Charges at 14.9% of Total Capital Investment			3,385,600
Overhead			
Plant, 20% of Conversion Costs			321,000
Administrative, 10% of Operating Labor			15,900
SUBTOTAL INDIRECT COSTS			3,722,500
Annual Cost for Sodium Disposal at $6.00/ton			496,100
TOTAL ANNUAL OPERATING COSTS			9,406,800

COAL ANALYSIS: % S = 1.5 HEATING VALUE = 10,500 TEMPERATURE = 400°F

	Annual Quantity	Unit Cost, $	Total Annual Cost, $
Direct Costs			
Delivered Raw Material			
Nahcolite	225,050 tons	32.50/ ton	7,314,100
Conversion Costs			
Operating Labor and Supervision	18,000 man-hr.	8.81/ man-hr.	158,600
Utilities			
Electricity	45,205,899	0.018/ kWh	813,700
Maintenance			
Labor and Materials			713,600
Analyses			49,400
SUBTOTAL DIRECT COSTS.			9,049,400
Indirect Costs			
Average Capital Charges at 14.9% of Total Capital Investment			4,870,900
Overhead			
Plant, 20% of Conversion Costs			347,100
Administrative, 10% of Operating Labor			15,900
SUBTOTAL INDIRECT COSTS.			5,233,900
Annual Cost for Sodium Disposal at $6.00/ton			1,012,700
TOTAL ANNUAL OPERATING COSTS			15,296,000

COAL ANALYSIS: % S = 2.0 HEATING VALUE = 10,500 TEMPERATURE = 400°F

	Annual Quantity	Unit Cost, $	Total Annual Cost, $
Direct Costs			
Delivered Raw Material			
Nahcolite	350,000 tons	32.50/ ton	11,375,000
Conversion Costs			
Operating Labor and Supervision	18,000/ man-hr.	8.81/ man-hr.	158,600
Utilities			
Electricity	53,411,060	0.018/ kWh	961,400
Maintenance			
Labor and Materials			713,600
Analyses			49,400
SUBTOTAL DIRECT COSTS.			13,258,000
Indirect Costs			
Average Capital Charges at 14.9% of Total Capital Investment			6,419,300
Overhead			
Plant, 20% of Conversion Costs			376,600
Administrative, 10% of Operating Labor			15,900
SUBTOTAL INDIRECT COSTS.			6,811,800
Annual Cost for Sodium Disposal at $6.00/ton			1,575,000
TOTAL ANNUAL OPERATING COSTS			20,227,300

COAL ANALYSIS: % S = 0.5 HEATING VALUE = 10,500 TEMPERATURE = 525°F

	Annual Quantity	Unit Cost, $	Total Annual Cost, $
Direct Costs			
Delivered Raw Material			
Nahcolite	21,350 tons	32.50/ ton	693,900
Conversion Costs			
Operating Labor and Supervision	18,000 man-hr.	8.81/ man-hr.	158,600
Utilities			
Electricity	37,345,252	0.018/ kWh	672,200
Maintenance			
Labor and Materials			818,300
Analyses			49,400
SUBTOTAL DIRECT COSTS.			2,392,400
Indirect Costs			
Average Capital Charges at 14.9% of Total Capital Investment			2,540,600
Overhead			
Plant, 20% of Conversion Costs			339,700
Administrative, 10% of Operating Labor			15,900
SUBTOTAL INDIRECT COSTS.			2,896,200
Annual Cost for Sodium Disposal at $6.00/ton			96,100
TOTAL ANNUAL OPERATING COSTS			5,384,700

COAL ANALYSIS: % S = 1.0 HEATING VALUE = 10,500 TEMPERATURE = 525°F

	Annual Quantity	Unit Cost, $	Total Annual Cost, $
Direct Costs			
Delivered Raw Material			
Nahcolite	75,250 tons	32.50/ ton	2,445,600
Conversion Costs			
Operating Labor and Supervision	18,000/ man-hr.	8.81/ man-hr.	158,600
Utilities			
Electricity	40,296,583	0.018/ kWh	725,300
Maintenance			
Labor and Materials			218,300
Analyses			49,400
SUBTOTAL DIRECT COSTS.			4,197,200
Indirect Costs			
Average Capital Charges at 14.9% of Total Capital Investment			3,200,500
Overhead			
Plant, 20% of Conversion Costs			350,300
Administrative, 10% of Operating Labor			15,900
SUBTOTAL INDIRECT COSTS.			3,566,700
Annual Cost for Sodium Disposal at $6.00/ton			338,600
TOTAL ANNUAL OPERATING COSTS			8,102,500

COAL ANALYSIS: % S = 1.5 HEATING VALUE = 10,500 TEMPERATURE = 525°F

	Annual Quantity	Unit Cost, $	Total Annual Cost, $
Direct Costs			
Delivered Raw Material			
Nahcolite	128,450 tons	32.50/ ton	4,174,600
Conversion Costs			
Operating Labor and Supervision	18,000/ man-hr.	8.81/ man-hr.	158,600
Utilities			
Electricity	43,489,074	0.018/ kWh	782,800
Maintenance			
Labor and Materials			818,300
Analyses			49,400
SUBTOTAL DIRECT COSTS.			5,983,700
Indirect Costs			
Average Capital Charges at 14.9% of Total Capital Investment			3,855,800
Overhead			
Plant, 20% of Conversion Costs			361,800
Administrative, 10% of Operating Labor			15,900
SUBTOTAL INDIRECT COSTS.			4,233,500
Annual Cost for Sodium Disposal at $6.00/ton			578,000
TOTAL ANNUAL OPERATING COSTS			10,795,200

COAL ANALYSIS: % S = 2.0 HEATING VALUE = 10,500 TEMPERATURE = 525°F

	Annual Quantity	Unit Cost, $	Total Annual Cost, $
Direct Costs			
Delivered Raw Material			
Nahcolite	182,000 tons	32.50/ ton	5,915,000
Conversion Costs			
Operating Labor and Supervision	18,000/ man-hr.	8.81/ man-hr.	158,600
Utilities			
Electricity	46,842,338	0.018/ kWh	843,200
Maintenance			
Labor and Materials			818,300
Analyses			49,400
SUBTOTAL DIRECT COSTS.			7,784,500
Indirect Costs			
Average Capital Charges at 14.9% of Total Capital Investment			4,517,100
Overhead			
Plant, 20% of Conversion Costs			373,900
Administrative, 10% of Operating Labor			15,900
SUBTOTAL INDIRECT COSTS.			4,906,900
Annual Cost for Sodium Disposal at $6.00/ton			819,000
TOTAL ANNUAL OPERATING COSTS			13,510,400

COAL ANALYSIS: % S = 1.0 HEATING VALUE = 8,000 TEMPERATURE = 275°F

	Annual Quantity	Unit Cost, $	Total Annual Cost, $
Direct Costs			
Delivered Raw Material			
Nahcolite	242,200 tons	32.50/ ton	7,871,500
Conversion Costs			
Operating Labor and Supervision	18,000/ man-hr.	8.81/ man-hr.	158,600
Utilities			
Electricity	49,983,152	0.018/ kWh	899,700
Maintenance			
Labor and Materials			799,200
Analyses			49,400
SUBTOTAL DIRECT COSTS.			9,778,400
Indirect Costs			
Average Capital Charges at 14.9% of Total Capital Investment			5,249,400
Overhead			
Plant, 20% of Conversion Costs			381,400
Administrative, 10% of Operating Labor			15,900
SUBTOTAL INDIRECT COSTS.			5,646,700
Annual Cost for Sodium Disposal at $6.00/ton			1,089,900
TOTAL ANNUAL OPERATING COSTS			16,515,000

COAL ANALYSIS: % S = 1.0 HEATING VALUE = 13,000 TEMPERATURE = 275°F

	Annual Quantity	Unit Cost, $	Total Annual Cost, $
Direct Costs			
Delivered Raw Material			
Nahcolite	102,900 tons	32.50/ ton	3,344,300
Conversion Costs			
Operating Labor and Supervision	18,000/ man-hr.	8.81/ man-hr.	158,600
Utilities			
Electricity	28,152,486	0.018/ kWh	506,700
Maintenance			
Labor and Materials			494,800
Analyses			49,400
SUBTOTAL DIRECT COSTS.			4,553,800
Indirect Costs			
Average Capital Charges at 14.9% of Total Capital Investment			2,911,700
Overhead			
Plant, 20% of Conversion Costs			241,900
Administrative, 10% of Operating Labor			15,900
SUBTOTAL INDIRECT COSTS.			3,169,500
Annual Cost for Sodium Disposal at $6.00/ton			463,100
TOTAL ANNUAL OPERATING COSTS			8,186,400

COAL ANALYSIS: % S = 1.0 HEATING VALUE = 8,000 TEMPERATURE = 400°F

	Annual Quantity	Unit Cost, $	Total Annual Cost, $
Direct Costs			
Delivered Raw Material			
Nahcolite	184,100 tons	32.50/ ton	5,983,300
Conversion Costs			
Operating Labor and Supervision	18,000/ man-hr.	8.81/ man-hr.	158,600
Utilities			
Electricity	52,055,944	0.018/ kWh	937,000
Maintenance			
Labor and Materials			937,200
Analyses			49,400
SUBTOTAL DIRECT COSTS.			8,065,500
Indirect Costs			
Average Capital Charges at 14.9% of Total Capital Investment			4,741,500
Overhead			
Plant, 20% of Conversion Costs			416,400
Administrative, 10% of Operating Labor			15,900
SUBTOTAL INDIRECT COSTS.			5,173,800
Annual Cost for Sodium Disposal at $6.00/ton			828,500
TOTAL ANNUAL OPERATING COSTS			14,067,800

COAL ANALYSIS: % S = 1.0 HEATING VALUE = 13,000 TEMPERATURE = 400°F

	Annual Quantity	Unit Cost, $	Total Annual Cost, $
Direct Costs			
Delivered Raw Material			
Nahcolite	68,950 tons	32.50/ ton	2,240,900
Conversion Costs			
Operating Labor and Supervision	18,000/ man-hr.	8.81/ man-hr.	158,600
Utilities			
Electricity	29,582,217	0.018/ kWh	532,500
Maintenance			
Labor and Materials			575,600
Analyses			49,400
SUBTOTAL DIRECT COSTS.			3,557,000
Indirect Costs			
Average Capital Charges at 14.9% of Total Capital Investment			2,665,800
Overhead			
Plant, 20% of Conversion Costs			263,200
Administrative, 10% of Operating Labor			15,900
SUBTOTAL INDIRECT COSTS.			2,944,900
Annual Cost for Sodium Disposal at $6.00/ton			310,300
TOTAL ANNUAL OPERATING COSTS			6,812,200

COAL ANALYSIS: % S = 1.0 HEATING VALUE = 8,000 TEMPERATURE = 525°F

	Annual Quantity	Unit Cost, $	Total Annual Cost, $
Direct Costs			
Delivered Raw Material			
Nahcolite	111,650 tons	32.50/ ton	3,628,600
Conversion Costs			
Operating Labor and Supervision	18,000/ man-hr.	8.81/ man-hr.	158,600
Utilities			
Electricity	53,210,094	0.018/ kWh	957,800
Maintenance			
Labor and Materials			1,070,400
Analyses			49,400
SUBTOTAL DIRECT COSTS.			5,864,800
Indirect Costs			
Average Capital Charges at 14.9% of Total Capital Investment			2,653,500
Overhead			
Plant, 20% of Conversion Costs			447,200
Administrative, 10% of Operating Labor			15,900
SUBTOTAL INDIRECT COSTS.			3,116,600
Annual Cost for Sodium Disposal at $6.00/ton			502,400
TOTAL ANNUAL OPERATING COSTS			9,483,800

COAL ANALYSIS: % S = 1.0 HEATING VALUE = 13,000 TEMPERATURE = 525°F

	Annual Quantity	Unit Cost, $	Total Annual Cost, $
Direct Costs			
Delivered Raw Material			
Nahcolite	54,600 tons	32.50/ ton	1,774,500
Conversion Costs			
Operating Labor and Supervision	18,000/ man-hr.	8.81/ man-hr.	158,600
Utilities			
Electricity	32,453,162	0.018/ kWh	584,200
Maintenance			
Labor and Materials			661,300
Analyses			49,400
SUBTOTAL DIRECT COSTS.			3,228,000
Indirect Costs			
Average Capital Charges at 14.9% of Total Capital Investment			4,068,300
Overhead			
Plant, 20% of Conversion Costs			290,700
Administrative, 10% of Operating Labor			15,900
SUBTOTAL INDIRECT COSTS.			4,374,900
Annual Cost for Sodium Disposal at $6.00/ton			245,700
TOTAL ANNUAL OPERATING COSTS			7,848,600

APPENDIX D

MAGNESIUM OXIDE AS A BAGHOUSE SORBENT

CONCEPT

Magnesium oxide powder could, based on its theoretical reactions with gaseous SO_2, be used as a sorbent material in conjunction with a baghouse FGD installation. Magnesium oxide is, however, too expensive to be considered as a feed material to a "throwaway" process and can only demonstrate practical economics when used in a regenerable scheme. The regeneration of magnesium sulfate and sulfite to the active oxide form is accomplished by calcination and a dry product is obtained. In the currently-developed magnesium oxide wet FGD process, the regenerated magnesium oxide must be slurried with water, used in the wet scrubber, and the spent magnesium salts must then be separated by centrifuge and dried prior to calcination. The use of a baghouse would eliminate these "wet" steps and holds considerable promise. The feasibility and economic practicality rests on two questions:

- Is magnesium oxide sufficiently reactive to obtain reasonable SO_2 removal efficiency when used in a baghouse at temperatures compatible with bag fabrics?
- Is a two stage baghouse system less expensive to build and operate than the "wet" portions of the current magnesium oxide FGD process (including particle control)?

The first question must, of course, be answered affirmatively before the economic analysis required to answer the second is warranted.

DISCUSSION

The thermochemical feasibility of using magnesium oxide (MgO) as a dry sorbent in a flue gas desulfurization scheme utilizing a baghouse as a collection medium was undertaken. This means of sulfur dioxide removal showed interest because MgO has the potential for not only absorbing SO_2 in the temperature range of interest (that of a typical flue gas from a pulverized coal utility boiler) but also is capable of being regenerated from the spent sorbent by means of thermal decomposition. In order to analyze the use of MgO in the dry additive/baghouse flue gas desulfurization system, it was necessary only to examine the chemistry of the absorbing section since the regeneration process has been well established through demonstration.

In the absorption step, a two baghouse design would be necessary in order to avoid collection of fly ash on the absorbing filter cake, thereby decreasing its efficiency due to dilution effects and possible blinding of the sorbing surface. More important, however, is the fact that the "all dry" magnesium oxide regeneration scheme does not lend itself to separation of collected fly ash and the sorbent.

The first baghouse would be used to collect the fly ash contained in the flue gas while MgO would be injected between the first and second baghouse.

Theoretically, SO_2 is absorbed by the MgO in accordance with the following reactions(1,2):

1) $MgO + SO_2 = MgSO_3$

$SO_2 + 1/2\ O_2 = SO_3$

2) $MgO + SO_3 = MgSO_4$

Figure D-1(1) illustrates the dependence of the reaction equilibrium of equation (1) on temperature. At a temperature of approximately 350°F (180°C) the flue gas SO_2 will combine with the MgO to reach an equilibrium sufficient to meet the New Source Performance Standards of 1.2 lb $SO_2/10^6$ Btu. At temperatures below this the reaction will favor increased SO_2 removal and vice versa. Thermal regeneration is possible at temperatures above 660°F (350°C) and improves with increasing temperature.

Figure D-2(1) illustrates the corresponding relationship for equation (2). The combination of MgO with SO_3 to meet NSPS levels is favored at temperatures at or below 1500°F (820°C). This reaction is not reversible at practical conditions.

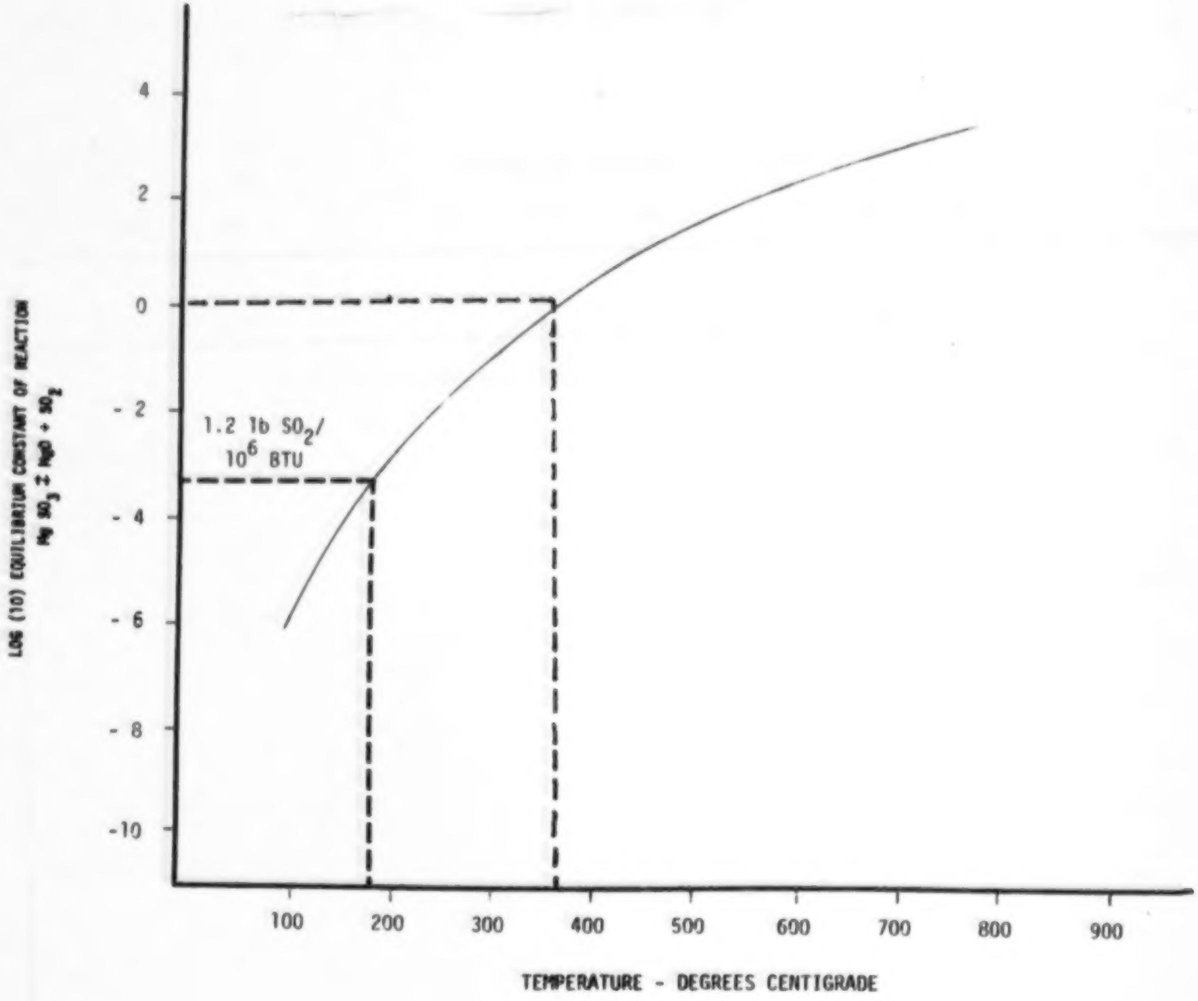


FIGURE D-1(1). $MgSO_3 \rightleftarrows MgO + SO_2$

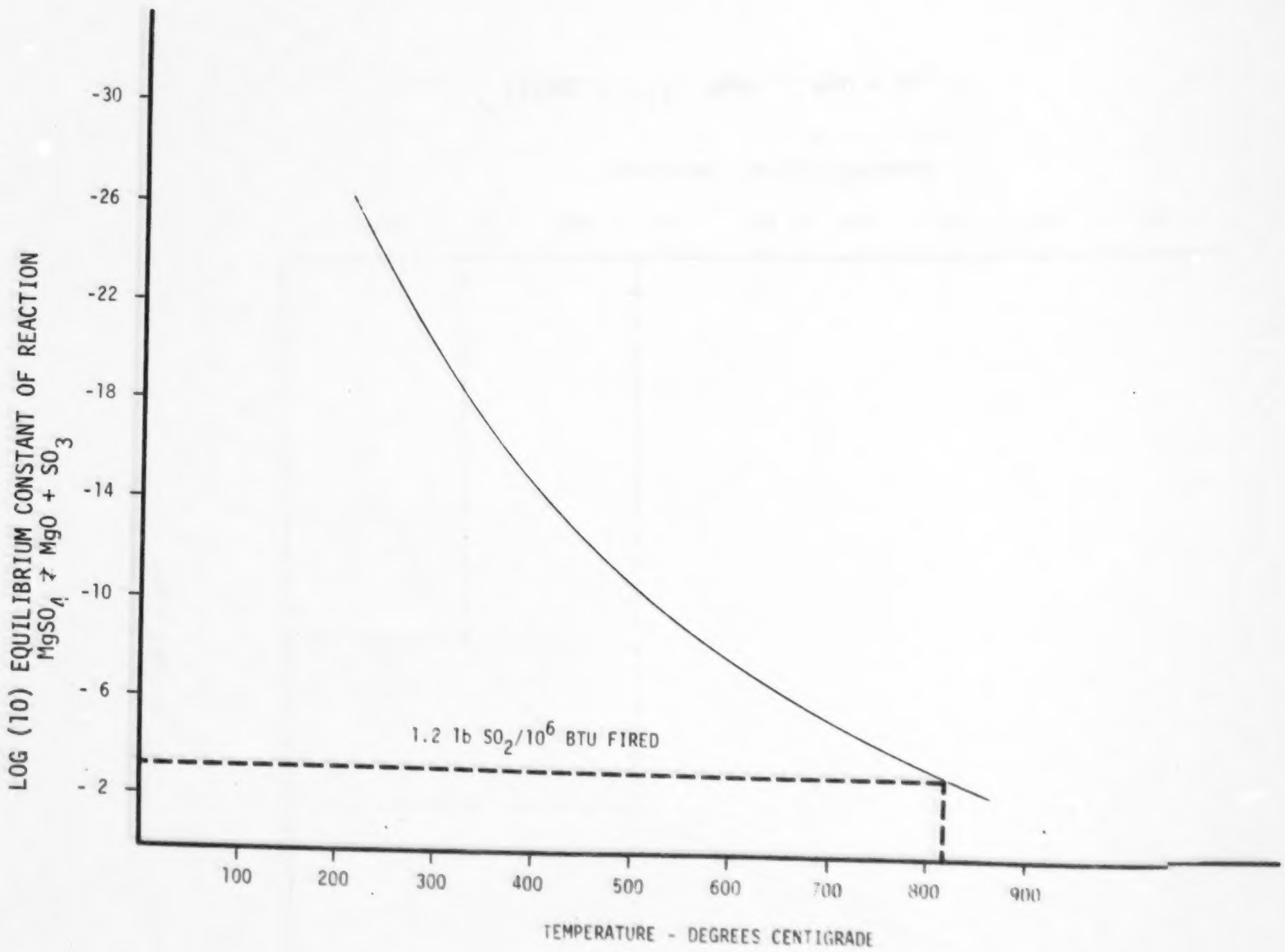


FIGURE D-2(1). $MgSO_4 \rightleftarrows MgO + SO_3$

From a thermodynamic standpoint, SO_2 capture by MgO would seem to be quite favorable in the temperature range associated with boilers. Nonetheless, operating within the thermodynamically-favored temperature range does not guarantee sufficient reductions in the SO_2 level since the reaction rates of the various ongoing reactions dictate the chemistry of the system. That is, kinetic considerations may override otherwise favorable thermodynamics.

To facilitate the determination of the multitude of possible reaction rates at which the SO_2 could interact with MgO, an upper bound of 830°C was chosen (which is the thermodynamic constraint put on the system by the NSPS for SO_2) while a lower limit of 135°C (which gives the optimum recovery of heat in the process from an economic point of view) was imposed. Extensive research has been done by Pechkovskii(4) and Marier and Dibbs(5) on the absorption of SO_2 by MgO in the temperature range of 600-900°C. Certain points should be noted, however, in inferring any conclusions from their work about the uptake of SO_2 in the flue gas desulfurization scheme under question. In his determination of SO_2 reaction rates with MgO, Pechkovskii used a gas atmosphere which contained 2.5 percent SO_2 and 97.5 percent air (volume percents) which has a ten times greater percentage of SO_2 by volume than the typical flue gas stream from a pulverized coal power plant. This, however, would tend to increase the percent conversion of MgO since the reaction rate is also dependent upon the concentrations of the reacting constituents. Also, Pechkovskii ran his experiments for a 15 minute period, which is two orders of magnitude (10^2) greater than the contact times envisioned for a dry sorbent/baghouse flue gas desulfurization process. Again, percent conversions of MgO deduced from Pechkovskii's work would tend to be greater than those expected from the MgO/baghouse operation.

Both Pechkovskii and Marier and Dibbs have demonstrated the catalyzing effects of Fe_2O_3 on the oxidation of SO_2 to SO_3. Though Fe_2O_3 is a major constituent of fly ash, its effect on the SO_2 in the flue gas is expected to be negligible due to the unreactive state of the fly ash. Due to the high temperatures in the boiler, the particulate matter tends to fuse, thereby decreasing the surface area for reaction (both on the surface of the particle and in the pore structure); this results in an appreciable decrease in the catalytic properties of the Fe_2O_3 in the fly ash. For this reason, SO_2 capture by MgO in the baghouse would be predominantly by non-catalyzed reactions.

Extrapolation of reaction rates of SO_2 uptake by Mgo from an Arrhenius plot given in Figure F-3 (data from which derived given in Table F-1) gives the following results:

Temperature	Percent Conversion by Weight of MgO*
135°C	0.03%
830°C	3.65%

*Percent conversion in a 15 minute run.

These two percent conversions bracket the expected conversion rate (which is determined by the temperature chosen to optimize the system). Since the greatest utilization rate of MgO in a one-pass system would be on the order of 4 percent and such a conversion is obtained only at temperatures well above baghouse operating temperature limits, it is obvious that to obtain sufficient utilization rates of MgO, a multi-pass recycle stream would have to be incorporated. The low uptake per pass would result in unreasonably large recycle rates and thus render the overall scheme unworkable.

TABLE D-1. TEMPERATURE DEPENDENCE OF MgO CONVERSION

Temperature °C	Percent Conversion of MgO (Weight Percent)
600	1.7
700	2.5
800	3.4
900	4.3

Gas Composition: 2.5% (by weight) SO_2
97.5% (by weight) Air

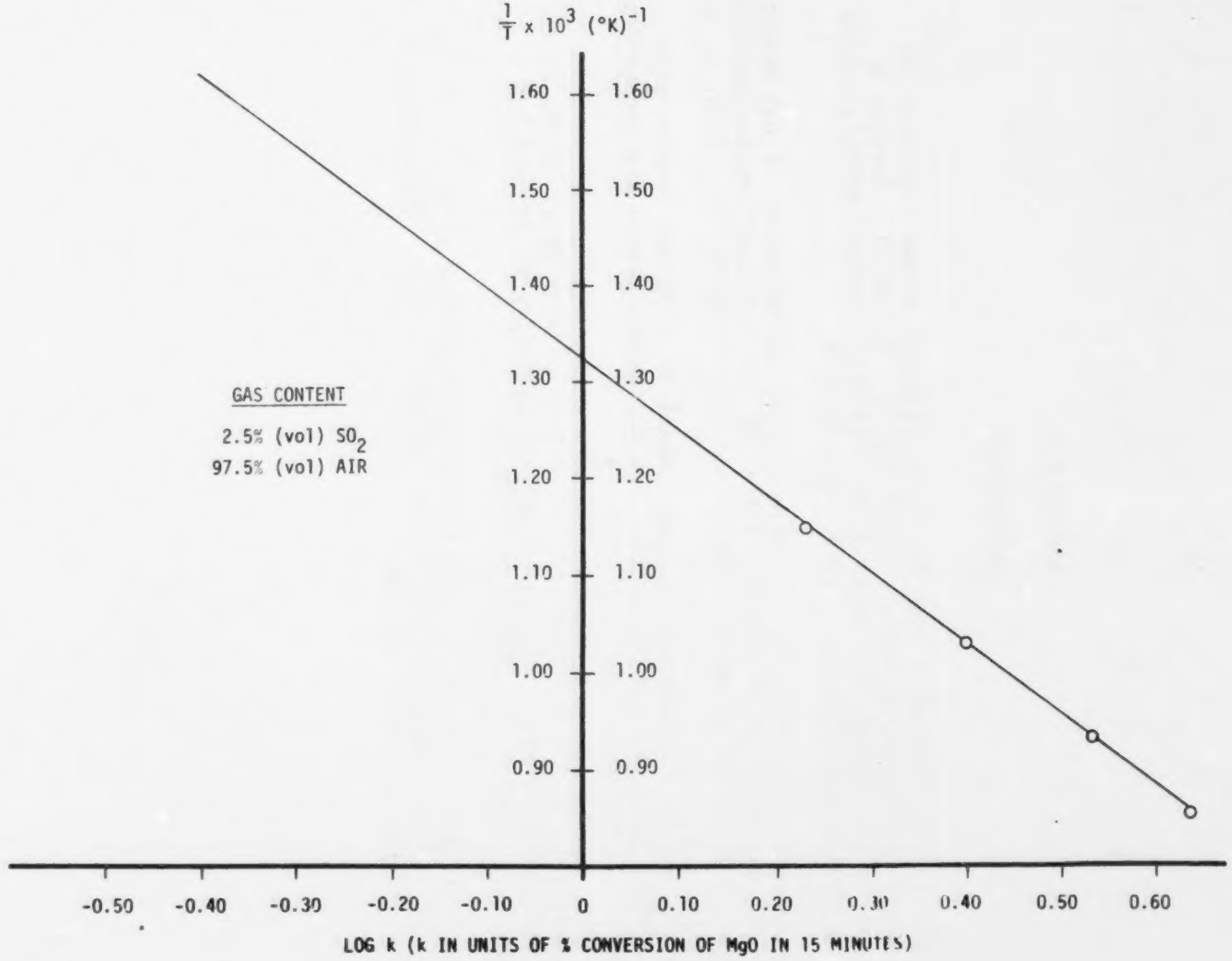


FIGURE D-3. ARRHENIUS PLOT OF SO_2 UPTAKE WITH MgO

APPENDIX D

REFERENCES

1. Applicability of Metal Oxides to the Development of New Processes for Removing SO_2 from Flue Gases. Final Report. Volume II. Section 8. Report done by Tracor for National Air Pollution Control Administration under contract PH 86-68-68. 1969. pp. 308-310.

2. Fundamental Study of Sulfur Fixation by Lime and Magnesia. Final Report. Report done by Battelle Memorial Institute for Taft Sanitary Engineering Center (Public Health Service) under contract PH 86-66-108. 1966. p. 6.

3. Murthi, K., D. Harrison and R. Chan. Reaction of Sulfur Dioxide with Calcined Limestones and Dolomites. Environmental Science and Technology. Volume 5. Number 9. 1971. p. 776.

4. Pechkovskii, V. V. Reactions of Sulfur Dioxide with Metal Oxides in an Oxidizing Atmosphere. Journal of Applied Chemistry of the U.S.S.R., Volume 30. Number 11. 1957. p. 1643.

5. Marier, P. and Dibbs, H. P. The Catalytic Conversion of SO_2 to SO_3 by Fly Ash and the Capture of SO_2 and SO_3 by CaO and MgO. Thermochemica Acta. Volume 8. 1974. p. 155.